SIMULATION OF AN INTEGRATED SYSTEM FOR THE PRODUCTION OF METHANE AND SINGLE CELL PROTEIN FROM BIOMASS

By

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To my wife, Laurie, without whose love and encouragement this would never have been accomplished.

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Abstract of Dissertation Presented to the Graduate School of the University of Florida in Partial Fulfillment of the Requirements for the Degree of Doctor of Philosophy

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A numerical model was developed to simulate the operation of an integrated system for the production of methane and single-cell algal protein from a variety of biomass energy crops or waste streams. Economic analysis was performed at the end of each simulation. The model was capable of assisting in the determination of design parameters by providing relative economic information for various strategies.

Three configurations of anaerobic reactors were simulated. These included fixed-bed reactors, conventional stirred tank reactors, and continuously expanding reactors. A generic anaerobic digestion process model, using lumped substrate parameters, was developed for use by type-specific reactor models. The generic anaerobic digestion model

provided a tool for the testing of conversion efficiencies and kinetic parameters for a wide range of substrate types and reactor designs.

Dynamic growth models were used to model the growth of algae and Eichornia crassipes as a function of daily incident radiation and temperature. The growth of Eichornia crassipes was modeled for the production of biomass as a substrate for digestion.

Computer simulations with the system model indicated that tropical or subtropical locations offered the most promise for a viable system. The availability of large quantities of digestible waste and low land prices were found to be desirable in order to take advantage of the economies of scale. Other simulations indicated that poultry and swine manure produced larger biogas yields than cattle manure.

The model was created in a modular fashion to allow for testing of a wide variety of unit operations. Coding was performed in the Pascal language for use on personal computers.

INTRODUCTION

The use of various agricultural wastes to produce energy and to feed livestock is not new. The anaerobic digestion of animal wastes to produce methane for domestic consumption has been practiced for over a hundred years (Meynell, 1978). However, due to the low cost of electricity and petroleum based fuels anaerobic digestion has not been widely accepted in the developed world as an energy source. With the growing awareness of our dwindling natural resources the potential use of biological methanogenic conversion processes is attracting increasing attention.

The economics of anaerobic production of methane in the developed world depend heavily on four main factors: the cost and availability of substrate; capital and operating costs of the plant itself; the costs associated with disposal of the waste stream; and revenues generated by the sale of products.

One of the methods to improve these economics is to combine related operations into an integrated system wherein a regenerative feedback is developed to maximize revenues while minimizing the associated costs. An operation of this

type studied at the University of Florida involved three unit operations located on a single site (Figure 1).

Swine waste and chopped water hyacinths were digested under anaerobic conditions to produce methane. Three types of digesters were used, a continuous stirred tank reactor (CSTR), a fixed-bed reactor (FBR), and a continuously expanding reactor (CER). Nutrient rich supernatant from the digesters was used to grow algae. The algae were harvested as a source of single cell protein for inclusion in animal feed formulations. Water leaving the algae unit flowed through shallow ponds filled with water hyacinths.

This research was an attempt to model an integrated system for methane and algal biomass production. The model was designed to be of a generic nature and was not specifically designed to represent the University of Florida system.

The objectives of this research were

1. To model an integrated system for use in the determination of proper design parameters.
2. To provide a research tool to assist in choosing values of conversion efficiencies and kinetic parameters for simulation of a variety of anaerobic digester configurations and substrates.
3. To simulate system performance as affected by feedstock characteristics, climate, harvest cycles, biogas usage, and flow management.

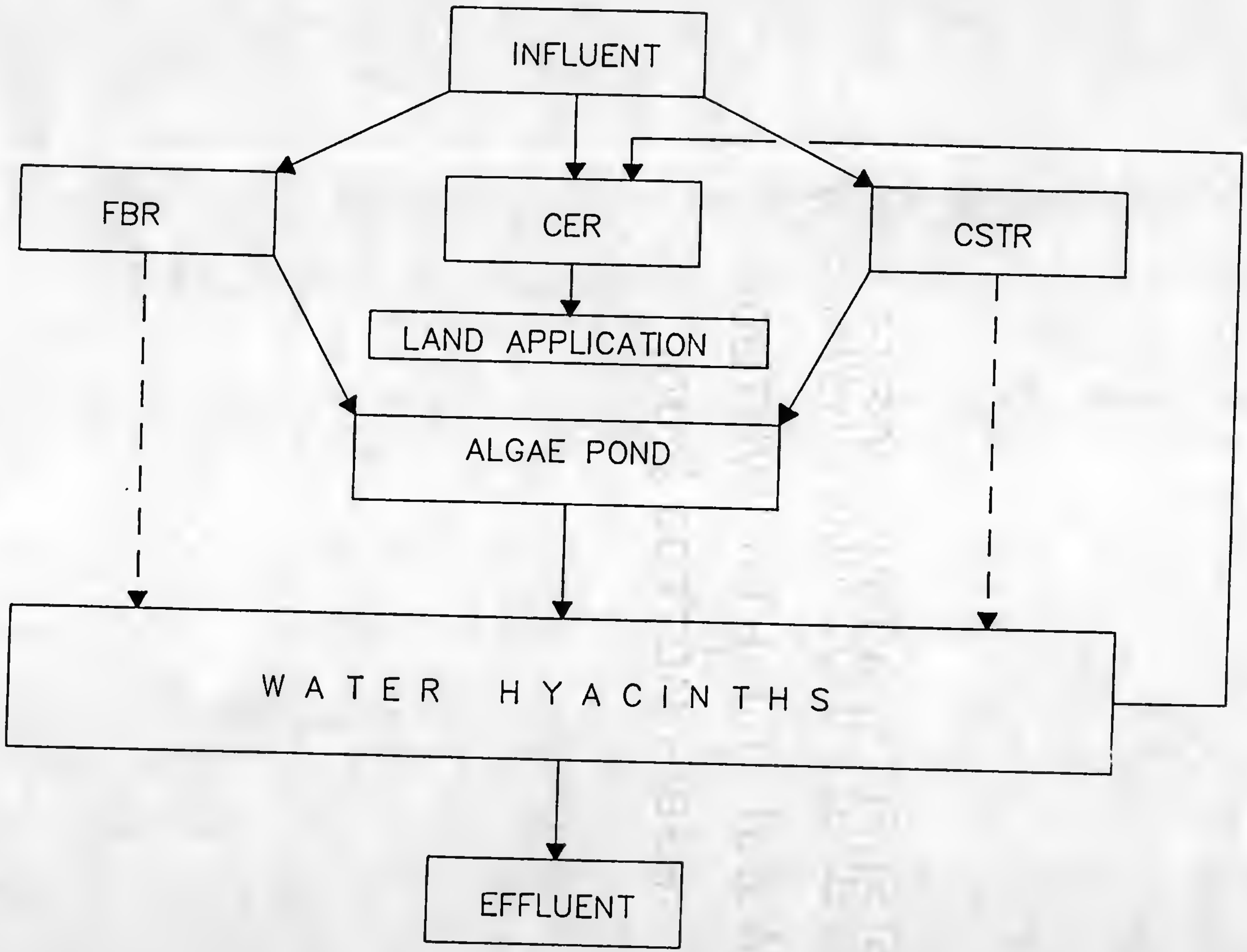

Figure 1. Simplified block diagram of the integrated system.

LITERATURE REVIEW

Unit Processes

Anaerobic Digestion

Anaerobic microbiology

Anaerobic digestion is a microbial process which has been widely used throughout the world for over a century to provide biogas for cooking and as a method of waste disposal. However, due to the continued availability of inexpensive petroleum based energy resources, biogas has not been used extensively in this country as an alternative energy source. Only in the last forty years has the microbiology of the process been studied. Much of the work has come since the oil embargo of the 1970's when the Organization of Petroleum Exporting Countries (OPEC) spurred the search for additional domestic energy sources.

The anaerobic process is particularly well suited for energy production. It is one of the most important parts of the carbon cycle since it results in the degradation of complex organic matter to relatively pure gaseous carbon dioxide (CO_2) and methane (CH_4) with a relatively small yield of bacterial mass (McCarty, 1964a,b). Thus, a large amount of organic matter is destroyed while about 90% of the

substrate energy is retained in the methane (Mah et al.,1976; Thauer, 1979; and Bryant, 1979).

The traditional view of the process classifies the organisms that carry out this process into two categories, the acetogenic bacteria and the methanogenic bacteria, as shown in Figure 2 (Barker, 1956; McCarty, 1964a). In this approach, the complex organic matter consisting of various proteins, carbohydrates and fat are broken down into simple short chain fatty acids and alcohols by the acetogenic, or acid-forming bacteria. These short chain fatty acids are metabolized by the methanogens into carbon dioxide and methane.

With the discovery in 1967 that Methanobacillus omelianskii (Barker, 1956; Wolfe et al., 1966) was in fact a symbiotic association of two bacteria (Bryant et al., 1967), it became clear that the methane pathway was much more complex than first believed. M. omelianskii was originally thought to metabolize ethanol and CO_2 to acetate and CH_4. However, it was later determined that one of the two organisms metabolized ethanol and water to acetate and H_2, while the other organism utilized the H_2 to reduce the CO_2 to CH_4. Other studies have shown that methanogens are unable to use alcohols other than methanol or to catabolize organic acids other than acetate and formate (Bryant, 1976,1979; McInerney et al., 1979). These discoveries, along with subsequent work by Mah et al. (1977) and

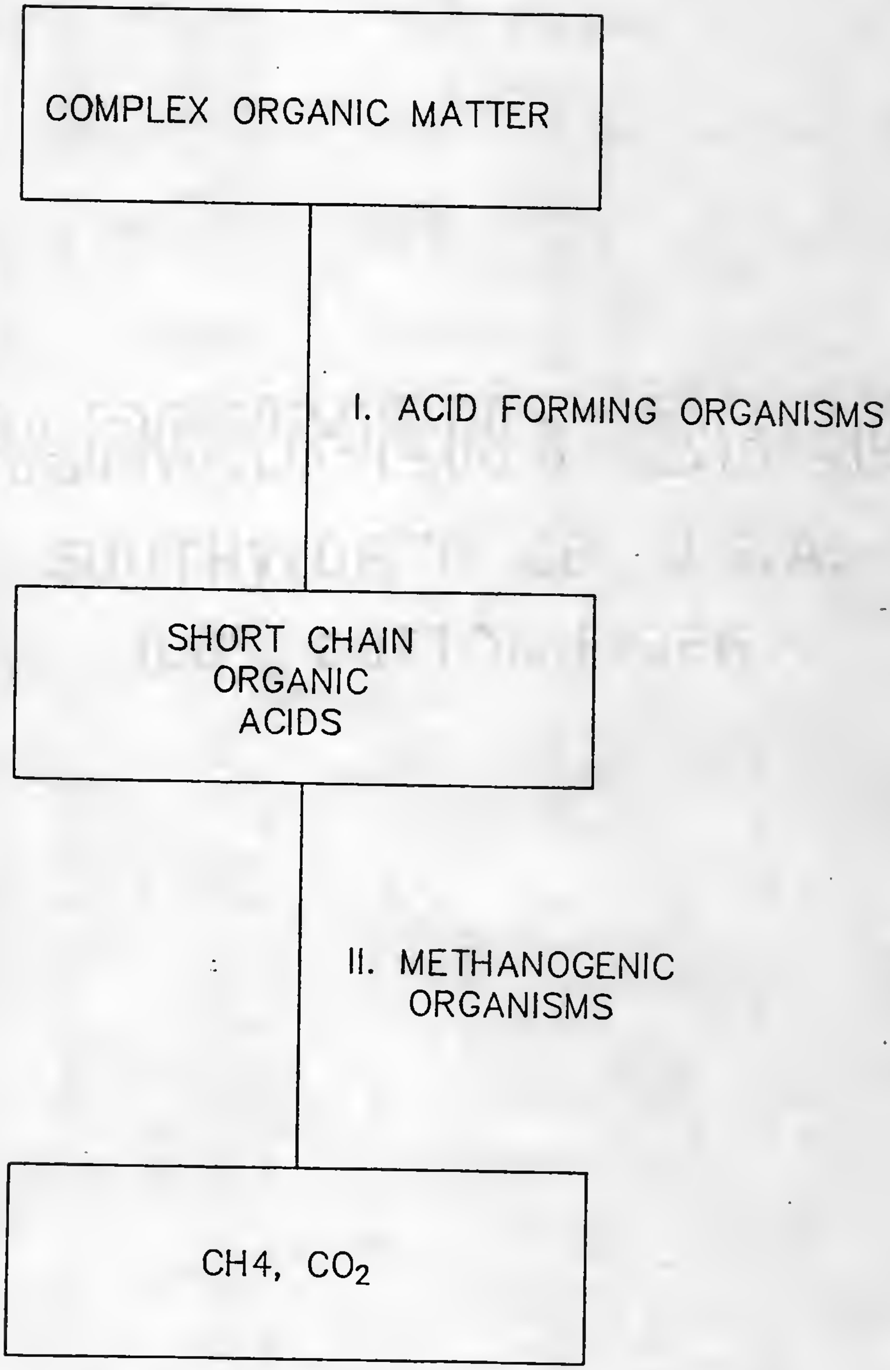

Figure 2. The two-population model of anaerobic digestion.

Thauer et al.(1977), led to the three stage model shown in Figure 3 (Bryant, 1979). In this model fermentative acid-formers produce short chain saturated fatty acids and alcohols. The H_2-producing acetogenic bacteria produce H_2 and acetate from the above end products, and the methanogenic bacteria produce CH_4 from acetate, CO_2, and H_2.

The critical importance of H_2 in the linking of this model cannot be overstated. Thauer et al. (1977) showed that the reactions which take place in the catabolism of propionate and butyrate to yield acetate and H_2 have a net positive change in Gibbs free energy. Only when these reactions are coupled to the strongly negative change in free energy associated with the reduction of HCO_3^- can the reaction proceed.

Wolfe and Higgins (1979), Zeikus (1979) and Hashimoto et al.(1980) described a fourth type of bacteria, the homoacetogens. This group uses H_2 to reduce CO_2 to acetate and is represented by such organisms as Acetobacterium and Clostridium aceticum.

The existence of an anaerobic fungus in the rumen of cattle and sheep was described by Bauchop (1979,1981). Fermentation of cellulose by this fungus, with subsequent methanogenesis by methanogenic bacteria in co-culture, has also been described (Bauchop and Mountfort, 1981; Mountfort et al., 1982). The four-population model (Figure 4) of anaerobic digestion represents our current understanding of

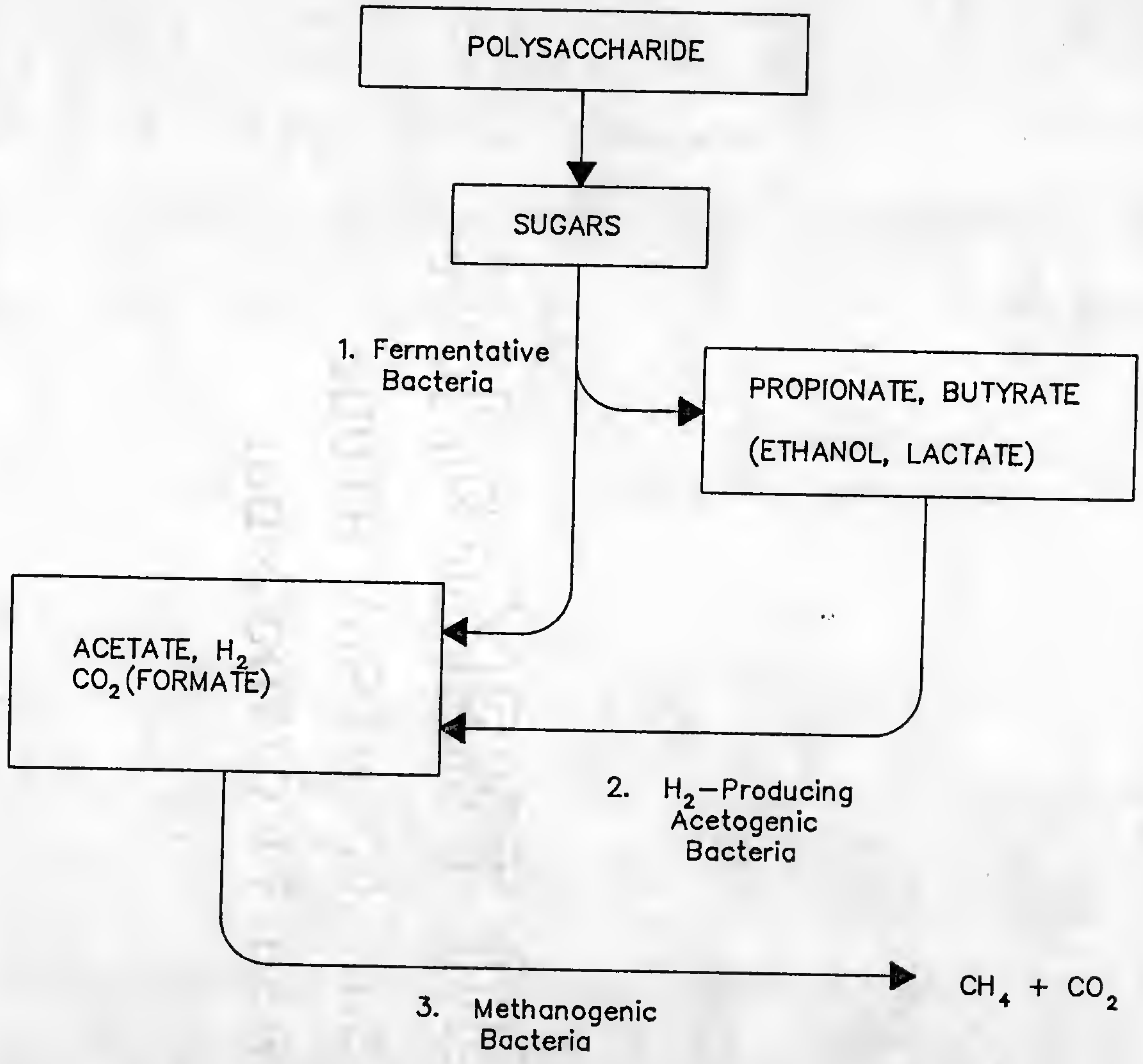

Figure 3. The three-population model of anaerobic digestion (Bryant, 1979). Reprinted by permission.

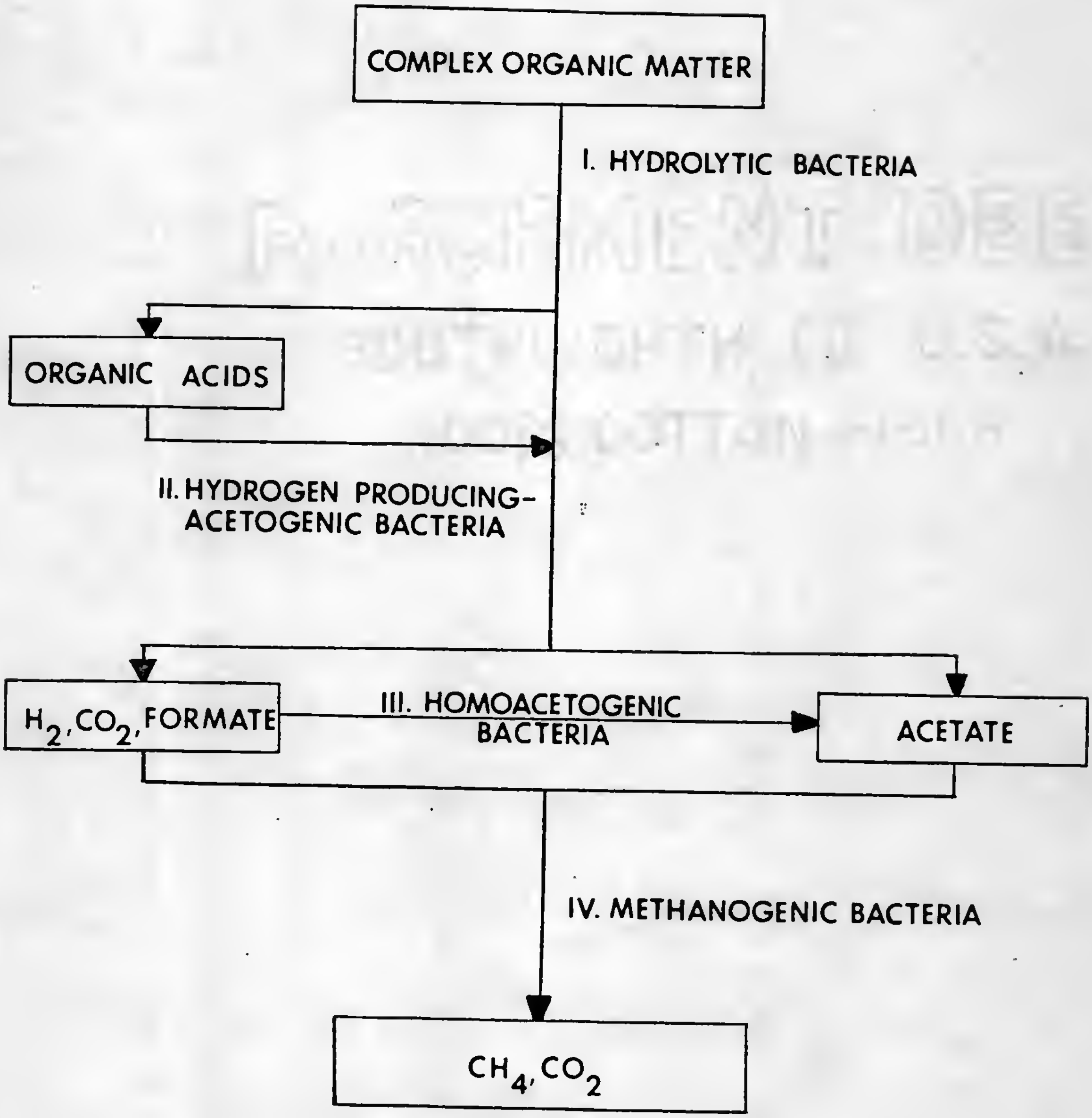

Figure 4. The four-population model of anaerobic digestion presented by Hashimoto et al. (1980). Reprinted by permission.

the anaerobic digestion process. The role of the anaerobic fungus is not well understood and no information was found in the literature on the subject of anaerobic fungal activity in digesters. As more information becomes available about the biochemistry of methane formation, particularly about the electron transfer reactions and role of molecular hydrogen in acetate catabolism, more pathways and populations will be defined.

Substrates for anaerobic digestion

Almost any nontoxic, biodegradable substance is a potential candidate for anaerobic digestion. The most readily digestible materials are obviously those soluble compounds which are intermediate products in the metabolic pathway. For example, the volatile fatty acids (VFA) may be used directly by H_2-producing acetogenic bacteria or methanogens. Simple sugars and amino acids that result from the action of the hydrolytic and proteolytic enzymes secreted by the acetogenic bacteria are quickly metabolized. The complex lipids, soluble carbohydrates and proteins are somewhat less readily degradable. These materials can be readily attacked and metabolized by enzymatic action. It has been shown that this group is responsible for most of the biogas production from a number of materials (Habig, 1985). Physical factors such as particle size and the

amount of surface area exposed to enzymatic activity may have a significant role in the rate of digestion.

Neutral and acid fiber materials such as cellulose, hemicellulose, and lignocellulose are much less readily digestible. Although cellulose and hemicellulose have been shown to be digestible (Khan, 1980; Laube and Martin, 1981; Singh et al., 1982; Nordstedt and Thomas, 1985a; Partos et al., 1982), the hydrolysis of this structural biomass is frequently the rate limiting step in the digestive process (Doyle et al., 1983; Chynoweth, 1987).

Lignin is not significantly degraded under most anaerobic conditions (Crawford, 1981; Singh et al., 1982; Nordstedt and Thomas, 1985b). Much of the emphasis on pretreatment of lignocellulosic biomass is an attempt to free the cellulose from lignin compounds and to make it available for digestion.

Historically human or animal manure, sometimes with the addition of crop residues, has been the substrate most frequently used. In the industrialized world the digestion of municipal sewage sludge has been a common practice. To a lesser extent, the digestion of animal manures has also been practiced.

Generally, the biodegradability (kg volatile solids (VS) destroyed/kg VS added) of manure from beef cattle is higher than that of dairy cattle due to the characteristics of the feed ration. The amount of bedding and other

extraneous material collected with the manure is also a factor and is influenced by the method of collection (Hill, 1983a; Loehr, 1984; and Chandler, 1980).

The biodegradability of manure from beef cattle in confinement was reported by Hill (1983a) to be about 0.65, whereas the factor for cattle on dirt feedlots was only 0.56. Hill also reported a biodegradability factor of 0.36 for dairy waste. Loehr (1984) reported ten values of biodegradability for dairy waste ranging from 0.37 to 0.62.

In an exhaustive study of substrate biodegradibility, Chandler (1980) reported a linear relationship between lignin content and biodegradability. The following relationship was established between biodegradability and lignin content:

$$B = -0.028X + 0.83,$$

where B is the biodegradable fraction, and X is the percent lignin of total volatile solids (sulfuric acid method).

Dairy manure is likely to contain a larger amount of bedding material and a higher lignin content than that from beef cattle. Loehr (1984) reported the lignin content of dairy manure to average 14.4 percent of dry matter, whereas the lignin content of beef manure was reported to be only 8.3 percent of dry matter. Jain et al. (1981) reported values of 9.9% and 12.7% lignin, respectively, as a percentage of dry matter following their analysis of sheep and cattle waste. Using the relationship developed by Chandler

(1980), B = 0.43 for dairy cattle, and B = 0.60 for beef cattle. These agree with the values reported by Hill (1983a) and Loehr (1984).

Poultry manure is substantially more biodegradable than cattle manure. Values as high as 0.87 (kg VS destroyed / kg VS added) have been reported for manure from layers, with somewhat lower values of 0.70 - 0.76 reported for broiler manure (Hill, 1983a; Loehr, 1984).

Swine waste is among the most suitable of all manures for anaerobic digestion. Biodegradability of swine waste was reported by Hill (1983a) to be 0.90. Swine waste also yields a higher percentage of methane in the biogas than most other manures (Loehr, 1984).

Methane productivity (m^3/kg VS added) was reported by Hill (1984a) as 0.32 for swine manure, 0.36 for poultry, and to vary between 0.13 - 0.24 for cattle waste.

In the last three decades a large number of other waste streams have been investigated. Many of these, particularly those in the food industry, are being successfully treated by anaerobic digestion. In many cases the methane is of secondary importance to the environmental effect, but it may provide a significant economic benefit.

Lovan and Foree (1972) reported on the treatment of brewery wastes, and Szendrey (1983) described a 13,000 m^3 anaerobic treatment system designed to produce methane from rum distillery wastes. The digestion of cheese whey, pear

peeling wastes, and bean blanching wastes was described by van den Berg et al. (1981). Stevens and van den Berg (1981) reported the use of tomato peeling wastes for digestion. Russell et al. (1985) announced an anaerobic sludge blanket reactor system for use in the treatment of potato processing wastes. Treatment of shellfish processing wastewater (Hudson et al., 1978), ethanol stillage wastes (Hills and Roberts, 1984), and tannery wastes (Cenni et al., 1982) have also been reported.

Since the OPEC oil embargo of the 1970's, attention has been focused on anaerobic digestion as a method of supplementing natural gas production from conventional sources. Although the use of manures and waste streams to produce methane may be of economic benefit to the farms or plants involved, the amount of biogas produced is rarely sufficient to justify off-site distribution and sale. In order to produce sufficient quantities of methane to supplement our commercial natural gas supplies it will be necessary to produce substrates on a large scale. This realization has led to an interest in the production of biomass crops which may be readily digested to methane in large quantities and on a continuous basis.

One of the potential crops that has been investigated at length in subtropical areas is the water hyacinth (Eichornia crassipes). Habig and Ryther (1984) investigated the methane yields from a number of substrates. They found

that water hyacinth produced 0.15 m^3/kg of volatile solids added and concluded that water hyacinths were a viable substrate for methane production. Much of the work with digestion of hyacinths has been done with hyacinths grown in the treatment of domestic sewage (Biljetina et al., 1987; Joglekar and Sonar, 1987). Blends of hyacinth and municipal sewage sludge are a more promising substrate for commercial methane production, a 2:1 blend yielding as high as 0.29 m^3 methane/kg VS added (Biljetina et al., 1985). Other potential substrates under investigation include napier grass (Pennisetum purpureum) and sorghum (Sorghum bicolor) (Chynoweth et al., 1984).

Anaerobic reactor configurations

Continuous stirred tank reactor. The classical design for continuous processing of wastes by anaerobic digestion is the continuous stirred tank reactor (CSTR) (McCarty, 1964a). It is basically a tank with inlet and outlet, and a motorized impeller for mixing. The mixing must be sufficient to prevent stratification and settling of the solids. It is assumed that the mixing is complete and that the solids residence time is equal to the hydraulic residence time, or RT = liquid volume/daily flow. Although this is the most popular design in current use, it suffers from several disadvantages.

The most obvious disadvantage is that it requires substantial mixing power (Hill, 1983b). In addition to the higher construction costs refecting the mechanical complexity required, the operating costs may also be significantly higher than for non-mixed reactors. Another disadvantage stems from the slow growth rate of the anaerobic organisms which define the process. The specific growth rate of many of the organisms involved is measured in days, instead of hours or minutes as is the case with most aerobic organisms. Thus, retention times for CSTR's are usually on the order of 10-30 days (McCarty, 1964a). Operation of CSTR's at less than 8-10 days retention time usually results in failure because the slower growing organisms wash out. The large capital investment required for a tank volume 15-30 times the daily flow makes the anaerobic CSTR economically infeasible in many situations. In an effort to minimize this size problem many CSTR's are operated at thermophilic temperatures of 50-55 C. The heating requirements are such that a large part of the methane produced be used for heating, although it can be minimized by heat recovery from the effluent (Hill 1983b) and use of waste engine heat from a generator. Thermophilic digesters also tend to be less stable than mesophilic digesters (Hill, 1983b; Hashimoto, 1983). In colder climates mesophilic digesters may require substantial external heating.

<u>Fixed bed reactor</u>. The fixed bed reactor (FBR), or anaerobic filter is designed to overcome some of the shortcomings of the CSTR when treating dilute waste streams. A porous packing material is used to provide a surface upon which bacterial growth can become attached. Since much of the bacterial mass is retained in the biofilm, bacterial washout is minimized. A large variety of packing materials and substrates have been utilized.

The first major investigation of fixed bed reactors was reported by Young and McCarty (1967). The columns were filled with smooth quartzite stones, 2.5 to 3.8 cm diameter, and were operated in an upflow mode. Four distinct advantages were listed for FBR's over conventional waste treatment systems:

1. FBR's are well suited to dilute wastes.
2. Biological solids are retained in the biofilm, allowing short hydraulic retention times.
3. High bacterial concentrations permit operation at nominal temperatures.
4. Very low volumes of sludge are produced.

Studies conducted on pharmaceutical waste by Jennett and Dennis (1975) used hand-graded smooth quartz stones 2.5 to 3.8 cm in diameter. Retention times varied between 12 and 48 hours with COD removal efficiencies of greater than 93% for all trials. Average methane concentration exceeded 80%.

Newell et al. (1979) reported on a 7.57 m^3 swine manure digester using 3.8 cm limestone chips as a support media. Gas quality exceeded 80% methane and COD removal averaged 90%.

Hudson et al. (1978) compared 2.5 to 3.8 cm. stone packing to whole oyster shells packing media for the treatment of shellfish processing wastewaters. The oyster shells were expected to provide buffering capacity in addition to providing a rough surface for biological attachment. Specific surface areas were approximately 130 m^2/m^3 for the stone and approximately 650 to 980 m^2/m^3 meter for the oyster shells. Gas production from the oyster shell digester was nearly double that of the digester with stone packing and the percent methane averaged over 85% as compared to only 73% with the stone packing.

Wilkie, Faherty, and Colleran (1983) examined the effects of media type upon performance of upflow anaerobic filters. Media which were examined included fired clay fragments, coral, mussel shells, and plastic rings. Maximum conversion of COD to methane was attained in 20 days for the clay fragments, and in 39, 40, and 50 days for the coral, plastic, and mussel shell filter media, respectively. Feed material consisted of swine manure slurry loaded at a rate of 5 kg COD/m^3 of liquid reactor volume per day at a hydraulic retention time of six days.

Person (1980, 1983) used polypropylene plastic media in his experiments with swine waste. The experiments were designed to compare the effectiveness of using media vs. no media, upflow vs. downflow, and to examine loading rates. Soluble COD removal was in excess of 90% and methane content was over 80% in those filters which contained media. Person (1980) concluded that filters with media are more effective than those without. It was also concluded that filters can operate with a hydraulic retention time as low as 9 hours at temperatures above 23.5 C and that most COD removal takes place in the lower 25% of the upflow filters.

Brumm (1980) examined the use of corncobs as a filter media in a series of experiments which compared filters packed with corncobs to those packed with plastic rings. The filters were loaded daily with dilute swine waste. The plastic media outperformed the corncobs in removal of influent volatile solids but lagged behind in the production of gas. This was determined to be due to the degradation of the corncob media. Methane content was also slightly lower for the corncobs, 74.3% vs. 78.3% for the plastic media.

Nordstedt and Thomas (1985b) conducted experiments using 13 bench-scale anaerobic filters packed with a variety of wood and plastic media. The wood media performed as well or better than the plastic media and showed no significant degradation after 1 year. Reactors were fed supernatant from settled swine waste. Methane content averaged

80% - 84%. Hydraulic retention times varied from 35 days at startup to a low of 2 days after 1 year of operation.

Continuously expanding reactor. The continuously expanding reactor (CER) represents another type of reactor. Unlike the CSTR and the FBR the CER does not operate in a continuous flow mode, but rather in a semi-batch mode. The substrate may be loaded into the reactor on a regular or irregular basis. However, the volume of the reactor is allowed to expand as it is loaded. It is only emptied on an infrequent basis, usually when it is convenient to apply the effluent to the land as a nutrient source. A "seed", or inoculum, consisting of 10% to 30% of the total digester volume is left behind after emptying to start the next cycle (Hill et al. 1981).

CER's tend to be more stable under conditions of very high loading than continuous processes and may provide a higher specific methane productivity (Hill et al. 1985). Most of the known work with CER's has involved the digestion of cattle manure (Hill et al. 1981) or swine manure (Hill et al. 1985).

Algae Production

The only method of solar energy conversion currently practiced on a large scale is photosynthesis. Unfortunately, the theoretical maximum efficiency for this process is

only about 5 or 6% (Hall, 1976). Microalgae, whose efficiency may be as high as 4% (Oswald, 1969), approach this theoretical limit to photosynthetic efficiency. The conversion efficiency rarely exceeds 1% in typical agricultural production. Even then the overall conversion efficiency is frequently negative as the caloric value of fertilizer and fossil fuels used exceed the value of the product (Benemann et al., 1976).

Algae have long been used in the treatment of organic wastes. In the conventional high rate pond the population is made up of about a 1:3 ratio of aerobic bacteria and algae (Oron et al., 1979; Hill and Lincoln, 1981). The aerobic bacteria stabilize the incoming waste and release CO_2 into the water. Algae utilize the CO_2 and sunlight to produce algal biomass through photosynthesis. Nitrogen composes about 10% of the dry weight of algal cells. When the cells are harvested and removed from the wastewater, up to 90% nitrogen removal may be obtained (Lincoln et al., 1977). Although total phosphates compose only 1-2% of the algal biomass, chemical flocculation of the algae for harvest usually results in excellent phosphate removal (Lincoln et al., 1980).

Because algae average about 50% protein considerable effort has been made to use algal protein as a supplement in human food and animal feeds (Yang, et al., 1981; Harrison, 1986; Lincoln and Earle, 1987). If problems of cell wall

digestibility can be overcome the feed value of the algae will far exceed the fuel value (Oswald, 1969; Lincoln et al., 1986). Algae are also known for a high lipid content. Algae may be economically used in some circumstances as a source of neutral lipids for commercial purposes (Dubinsky et al., 1978).

Species control must be achieved in an algal system in order to maintain a uniform product with consistent nutritional qualities. Flocculation and harvesting considerations also dictate that some type of species control be practiced. Benemann et al. (1976) showed that a degree of species control could be achieved with selective recycling. Lincoln and Earle (1987) reported that rotifers may be controlled by adjusting the pH and ammonium ion levels.

The high ash content associated with iron or aluminum flocculating agents may be overcome by the use of organic polymers and autoflotation (Koopman and Lincoln, 1983) or biological flocculation with gravity sedimentation (Koopman et al., 1987).

Water Hyacinth Production

Water hyacinths (*Eichornia crassipes*) are one of the most prolific aquatic plants found in tropical and subtropical regions of the world. They are usually regarded only as a weed, the state of Florida alone spending several million dollars per year for control (Bagnall, 1980). Given

adequate space and nutrients, a small mat of plants will double in area every 6 to 18 days. The mass is roughly proportional to the area. Wet densities are frequently in the range of 20 to 40 kg/m^2.

The water hyacinth has many characteristics that may make it an economically useful plant. It has a balanced nutrient content (Chynoweth et al., 1984) and is readily digestible to methane. It exhibits a daily growth rate of 20 to 40 g/m^2-d of dry biomass. Hyacinths may assimilate 10 kg N/ha-d or more, thus facilitating nitrogen removal from effluent streams. Nitrogen removal rates of 1726 to 7629 kg N/ha-yr were reported by Reddy et al. (1985). Several wastewater treatment processes in the United States currently use hyacinths (Stewart et al., 1987). Joglekar and Sonar (1987) determined that hyacinths could be used to treat up to 1250 m^3/ha-d of municipal wastewater while yielding 290 kg/ha-d of dry biomass.

DeBusk and Reddy (1987) determined that maintenance of high densities (1000 g/m^2 dw) maximized biomass yields. However, it has been shown that luxury uptake of nutrients is most pronounced when the plants are growing slowly (Reddy et al., 1985).

Because the dense stands of water hyacinths are free floating they are easily harvested (Bagnall, 1980). They can be chopped and fed into a digester as a slurry or they may be pressed to separate the solids from the juice.

Chynoweth et al. (1984) were successful in digesting the juice fraction separately from the solids. The juice was found to contain up to 25% of the biogasification potential of the entire plant.

Integrated Systems

An important consideration when examining energy production systems is to compare the cost and forms of energy produced with that of the energy consumed. The fundamental objective of developing biomass as an energy source is to capture solar energy in the form of plant tissue and to convert it to a higher quality form of energy (Jenkins and Knutson, 1984). To be truly useful, as opposed to economically advantageous only in the short term, the system must require less high quality energy than it produces. This requires that the system be studied using the concept of embodied energy in fossil fuel equivalents as the true measure of feasibility (Odum and Odum, 1976). However, since there is no demand for a system which is economically infeasible, the economic aspect should be examined first.

The successful integration of energy production processes into an agricultural or waste treatment system requires detailed information on how the various components will interact with all of the other components in the system. Material and energy flows must be evaluated, and logistical,

managerial, and environmental constraints must be identified (Walker, 1984).

Hayes et al. (1987) reported the operation of an integrated system for the treatment of municipal waste by water hyacinths. Methane production from digestion of hyacinths and sludge was a major goal of the project. It was determined that methane could be produced from such an operation at a cost of less than $2.00 /GJ in large cities.

Walker et al. (1984) analyzed operations at a large dairy farm in New York. It was determined that a combination of practices could reduce fossil fuel energy demands by 60%. Both energy conservation and fuel substitution were utilized. The utilization of the methane which was produced from wastes was a significant factor. The use of methane to fuel a cogeneration system was determined to be the best use of the gas.

Yang and Nagano (1985) investigated a system using an algal biomass raceway to provide additional treatment for anaerobically digested swine waste. The focus of the study was on the treatment aspects of the system and not on the production of algal protein.

Chen (1984) discussed sweet potato production from a systems point of view. The focus of this work was to determine optimal times and methods for planting.

Hill (1984b) optimized methane fermentation at swine production facilities. The parameter used for optimization

was the unit energy production cost. It was determined that the maximum economic return differed substantially from the point of maximum methane production. Heavier loading and shorter retention times were more favorable from an economic viewpoint.

Mathematical Modeling

Kinetics of Anaerobic Digestion

General kinetics

As the use of anaerobic digestion has increased in the last few decades so has the interest in the mathematical modeling of both digestion systems and the digestion process. Modeling of such systems may yield many benefits. In the process of designing a system, it is usually much faster and less expensive to model a system and simulate its operation on a digital computer than it is to actually build a series of pilot scale digesters and operate them over a long period. It also contributes to an understanding of the processes involved. Modeling forces the investigator to quantify the relationships between the component processes. It highlights the inconsistencies and tends to bring out the weaknesses in theoretical knowledge. In this manner, modeling tends to focus the direction of future experimental research.

Lawrence and McCarty (1969) recognized the need for a knowledge of the process kinetics of anaerobic digestion and

proposed the following model using the bacterial growth kinetics proposed by Monod (1949).

$$\frac{dM}{dt} = a\frac{dS}{dt} - bM$$

and

$$\frac{dS}{dt} = \frac{kMS}{K_s + S}$$

where

M = Concentration of microorganisms, mass/volume,

dM/dt = net growth rate of microorganisms,

dS/dt = rate of substrate assimilation,

a = growth yield coefficient,

b = microorganism decay coefficient,

S = substrate concentration, mass/volume,

k = maximum rate of substrate utilization per unit weight of microorganisms, $time^{-1}$,

K_s = half velocity coefficient for substrate utilization.

Combining these equations yields:

$$\frac{(dM/dt)}{M} = \frac{akS}{K_s + S} - b$$

The quantity (dM/dt)/M may be designated the net specific growth rate, mu. However, most later work defines mu as the specific growth rate and leaves the decay function to be examined separately. The maximum specific growth rate, mû,

is frequently used to represent a·k, yielding the more familiar form of the Monod equation:

$$mu = \frac{m\hat{u}\ S}{K_S + S}$$

The solutions to the above equations may be found for steady state conditions by setting the derivatives to zero and solving algebraically. However, transient behavior, which must be examined in order to model maximum system performance and digester failure, must be solved by numerical techniques.

Contois (1959) proposed the following modification to the Monod equation:

$$mu = \frac{m\hat{u}\ S}{BP + S}$$

where B is a constant growth parameter, and P is the bacterial density.

Monod defined K_S, the half velocity coefficient for nutrient utilization, as a constant for a given nutrient and bacterial population. Contois determined that K_S appeared to vary with population density. No microbiological basis has been postulated for this relationship. Contois postulated that the appearance of bacterial density in this relationship may be due to an inhibitory buildup of end products. This theory was born out by the work of Grady et

al. (1972), which showed that end product excretion varied with both the density and growth rate of the population. They proposed that until these relationships were better understood, especially in mixed cultures, the generation of statistical regression equations from operating data of similar reactors provided the most reliable model for design purposes. However, such equations are of little use in dynamic modeling of a system.

McCarty (1971) presented a series of possible stoichiometric reactions for methanogenesis in order to more accurately determine methane production. Reactions for both the catabolism of carbohydrate to methane, CO_2, and water and for the anabolic synthesis of bacterial cell mass were presented.

Andrews and Graef (1971) presented a dynamic model of anaerobic digestion using Monod kinetics. This model included the inhibitory effect of unionized fatty acids. This has been shown to be an important factor in predicting digester failure due to factors other than bacterial washout. The model also included a pH and alkalinity balance. This was necessary to determine the fraction of unionized fatty acids as well as to determine CO_2 gas transfer equilibria. The inhibitory effect of the fatty acid concentration was included in the microorganism specific growth rate by modifying the Monod equation as follows:

$$mu = \frac{\hat{mu}}{1 + \frac{K_s}{S} + \frac{S}{K_i}}$$

where K_s is the saturation constant, K_i is the inhibition constant, and S is the substrate concentration. The substrate concentration when the specific growth rate is at the maximum is then $S_m = (Ks \cdot Ki)^{0.5}$.

Hill and Barth (1977) provided a substantial expansion upon the work of Andrews and Graef. The maintenance of a nitrogen balance and the addition of an ammonia inhibition term to the methanogenic growth rate equation were of particular importance. The carbonate mass balance and pH calculations were also expanded.

The inhibitory effect of ammonia was included in the specific growth rate of methanogens by modifying the Andrews and Graef equation as follows:

$$mu = \frac{\hat{mu}}{1 + \frac{K_s}{S} + \frac{S}{K_i} + \frac{NH_3}{K_{iN}}}$$

Where K_{iN} is the inhibition constant for ammonia and NH_3 is the unionized ammonia concentration.

Hill and Nordstedt (1980) applied the work of Andrews and Graef (1971) and of Hill and Barth (1977) to anaerobic lagoons and anaerobic digesters. Yield coefficients and

mass transfer coefficients were determined for a number of parameters.

There are several compromises to be made in selecting which type of model to use. Simple first order models such as those proposed by Grady et al. (1972) and Srivastava et al. (1987) are capable of predicting steady state operation and require relatively few inputs. However, they are unable to predict process failure. Monod based dynamic models may very accurately predict process behavior, but, they require a very large number of kinetic parameters which are frequently unavailable and can only be estimated by computer iteration (Hill, 1983a).

A third type of model was developed in an attempt to bridge this gap (Hashimoto et al., 1980). This model was adapted from the kinetics of Contois for a completely mixed continuous flow system (Chen and Hashimoto, 1978; Hashimoto et al., 1980). The model predicts the volumetric methane production rate according to the following equation:

$$Y_v = \frac{Bo\ So}{\Theta} \left[1 - \frac{K}{\Theta\ m\hat{u} - 1 + K}\right]$$

where

Y_v = Steady state volumetric methane productivity L/L day

Bo = ultimate methane yield, L/gm VS added

So = influent VS concentration, gm/L

Θ = retention time, days

$\hat{mu}$ = maximum specific growth rate, day^{-1}

K = kinetic parameter, dimensionless.

The above equation states that for a given loading rate (So/θ) the volume of methane produced per day per liter of digester depends on the biodegradability (Bo) of the material and the kinetic parameters $\hat{mu}$ and K. The above equation is based upon the following (Chen and Hashimoto, 1978):

$$mu = \hat{mu} \left[\frac{\frac{S}{So}}{K + \frac{(1 - K) S}{So}} \right]$$

It should be noted again that this model was derived for steady state conditions. Under dynamic conditions it may be seen that mu approaches $\hat{mu}$ as S approaches So, regardless of the value of K. Therefore, this model will not predict process failure due to inhibition, although it is able to predict washout of the microbial population.

The Contois model of Chen and Hashimoto is adequate for many types of engineering analysis. It accurately reflects the methane productivity for steady state operation. Design and optimization of anaerobic digesters was extensively investigated by Hill (1982a, 1982b) using this model. Biodegradability (Bo) and K were both affected by the waste type, and K also varied with the VS loading rate. It was

determined that maximum volumetric methane productivity occurred under substantially different conditions than did the maximum daily production of methane.

Hill (1983a) recognized the value of having only a few parameters, as in the Chen and Hashimoto model, while retaining the advantages of non steady state kinetics. Hill used Monod kinetics in his "lumped parameter" model to obtain accurate dynamic characteristics. However, in order to simplify the model as much as possible many of the typical parameters were "lumped" into only two parameters which varied with the type of waste being treated. These parameters are the "biodegradability factor" (BO), and the "acid factor" (ACFACT). To avoid confusion, it should be noted that the biodegradability factor (BO) is the fraction of VS which is biodegradable, whereas the biodegradability (Bo) used by Hashimoto is the ultimate methane yield.

The assumption that all waste types can be reduced to homogeneous organic mixtures of biodegradable volatile solids (BVS) or volatile fatty acids (VFA) was fundamental to the derivation of this model. The amount of BVS and VFA was dependent upon the BO and ACFACT of the waste type. These parameters have been determined for a number of waste types (Hill, 1983a).

The model consisted of only six differential equations. Four of these were mass balances and two were microbial growth rate equations. Eight kinetic constants were

required for these six equations, along with two yield coefficients. However, it should be noted that these were true constants and did not vary with waste type.

Additional studies were conducted into the kinetics of microbial death (Hill et al., 1983; Hill, 1985). Continuously expanding reactors (CER's) were used in these studies in order to avoid interference from microbial washout. Most previous models fixed the specific death rate, K_d, at one tenth of the maximum specific growth rate. It was determined that this did not provide for adequate removal of viable microorganisms when there was no term for washout, as in a CER. A new death rate coeffcient was selected. It was assumed that the maximum death rate was not likely to exceed the maximum growth rate except under toxic conditions. Therefore, the following equations were proposed:

$$\hat{K}_d = m\hat{u}$$

and

$$K_d = \frac{\hat{K}_d \cdot VFA}{1 + K_{id} / VFA}$$

where K_{id} is the half velocity death constant for VFA.

A substantially more complex version of this model was presented by Hill (1982c). In this model the newly defined hydrogen-producing acetogenic bacteria and the homoacetogenic bacteria were included for the first time. For mathematical reasons, it was also necessary to simulate two

groups of methanogens, one using H_2-CO_2 as a substrate and the other using acetate. Five growth equations and twelve mass balances were required. The model was validated with previously published data.

Dwyer (1984) modified Hill's five population model by adjusting the stoichiometric relationships of the acetogenic reaction. This resulted in the release of additional hydrogen and led to more accurate predictions of gas quality. In addition, some of the kinetic parameters for the growth models of various populations were changed from the values used by Hill to yield a better fit to the available data.

Additional work has been done on modeling the transitory periods during digester start up and failure. Hill and Bolte (1987) reported significant changes in modeling of the hydrogenogenic population. Inhibition of this population was determined to be dependant on both ammonia and total VFA concentrations. Separate inhibition constants were derived for each substance. The uptake of propionate and butyrate was also modified. The proportional uptake concept was replaced by a competitive substrate concept. These changes made the five population model much more accurate in predicting propionic-acetic acid ratios, which was considered crucial to predictng digester failure (Hill et al., 1987).

CSTR modeling

Most of the kinetic equations developed above were designed to describe a continuous flow homogeneous reactor, or chemostat. This condition is most nearly approached by the CSTR. Transient variations in the bacterial population for such a reactor may be described as follows (Chen and Hashimoto, 1978):

$$\frac{dX}{dt} = \left(mu - \frac{1}{\theta_s} \right) X$$

and

$$\frac{dS}{dt} = \frac{S_o - S}{\theta_h} - \frac{mu \cdot X}{Y}$$

where,

X = bacterial concentration,

Y = bacterial yield coefficient.

θ_s = biologically active solids residence time.

θ_h = hydraulic residence time.

By definition, $\theta_s = \theta_h$ in an ideal CSTR where homogeneity is achieved.

FBR modeling

The fixed bed reactor, or anaerobic filter, presents a more complicated system than the CSTR. Because much of the biological activity in a FBR is associated with the biofilm, the residence time of the biological solids is much longer than the hydraulic residence time. Early attempts at

modeling of fixed film systems were made by Kornegay and Andrews (1967) and by Mueller and Mancini (1975).

Bolte et al. (1984) presented a complex model of an upflow anaerobic filter which portrayed the process dynamics associated with the biofilm. However, this model is very cumbersome to use because of the complex transport phenomena involved.

Another model for the FBR was proposed by Bolte and Hill (1985). This model was derived from the work of Hashimoto et al. (1980).

$$Y_v = \frac{Bo\ So}{\theta_h} \left[1 - \frac{K}{\theta_h\ m\hat{u} - \frac{\theta_h}{\theta_s} + K}\right]$$

In an attached growth reactor such as a fixed bed reactor, $\theta_s >> \theta_h$, so the equation reduces to :

$$Y_v = \frac{Bo\ So}{\theta_h} \left[1 - \frac{K}{\theta_h\ m\hat{u} + K}\right]$$

The model predicted steady state methane production within ten percent of the data values used for validation. However, it is only applicable to steady state conditions.

In an attempt to make the lumped parameter model of Hill (1983a) applicable to attached growth reactors, Bolte (1985) proposed the use of the "bacterial retention coefficient", or BRC. The BRC is a function of the type of reactor, recycle rate, surface to volume ratio, and media

type. The BRC represents the amount of active biomass which is retained in the reactor by modifying the washout term in the bacterial mass balance equation such that,

$$\frac{dX}{dt} = [mu - K_d - (1\text{-BRC}) \frac{1}{\theta_h}] \; X$$

Bolte did not validate his model and no additonal work in the literature was found which made use of the BRC to characterize retained biomass effects.

Another technique used in the modeling of nonhomogeneous reactors is the simulation of plug flow conditions by the use of several homogeneous reactors placed in series with one another. In theory this would require an infinite number of compartments of zero width. Two to ten compartments are commonly used in practical simulations (Bolte, 1985).

CER modeling

The continuously expanding reactor (CER), or semi-batch reactor, presents another challange. Because the volume is continuously changing, one must work with the total biomass present.

Young (1979) modeled a CER using beef cattle waste as a substrate. He used the kinetics which were developed by Hill and Barth (1977) and modified them to use extensive rather than intensive variables. The model performed reasonably well, although predictions of digester failure

tended to be protracted and to take place at lower than actual temperatures.

Hill et al. (1983) used a CER model to develop the new death kinetics reported earlier. This model used the lumped parameter model of Hill (1983a) with modified mass balances and extensive variables.

Algal Growth Kinetics

Most of the literature on the growth kinetics of algae deals with pure culture work in the laboratory. Very little information is available on mixed culture algal growth in the natural environment.

Enebo (1969) proposed the first order growth equation $dN/dt = k \cdot N$, where N is ammonia nitrogen and k is a function of species, temperature, substrate concentration, pH, illumination, etc. More recent work by Williams and Fisher (1985) on NH_4 uptake, by Gotham and Rhee (1981) on phosphate uptake, by Terry et al. (1985) on N:P ratios and by Turpin (1986) on C:P ratios are limited in their applicability to mixed systems where species composition is not constant.

A study on light limitation of growth rate by Schlesinger and Shuter (1981) provided a more general view. Their results on the effects of light limitation on the relative amounts of chlorophyll A and RNA were applicable to a wide range of species. Further research into the effects of light on protein synthesis could lead to more mechanistic

models of algal growth. When developed, these models may be applicable to a wider range of species than currently available models.

Hill and Lincoln (1981) assumed that microalgal kinetics were similar to bacterial kinetics and proposed the use of a modified Monod equation to represent algal growth in a mixed pond. Orthophosphate, ammonia nitrogen, carbon dioxide and incident solar radiation were treated as substrates which could be limiting. The specific growth rate for algae was determined to be:

$$mu_{alg} = \hat{mu}_{alg} \cdot (mu_{lim}) \cdot (1.05)^{T-25}$$

where mu_{lim} was equal to the mu_{sub} of the most severely limiting substrate as determined by the equation:

$$mu_{sub} = \frac{[Sub]}{K_{s_{sub}} + [Sub]}$$

where "sub" is one of the aforementioned substrates. The temperature dependancy function was an Arrhenius relationship based on 25 degrees C for each log cycle change in rate.

The model was calibrated using experimental data from the algal pond at the University of Florida Swine Research Unit in Gainesville, Florida. It was determined that the ratio of algal biomass to bacterial biomass was about 3:1. It was also determined that algae growing on the anaerobic

swine lagoon effluent were not nutrient limited and that light was the only limiting factor.

Bolte et al. (1986) found the model of Hill and Lincoln (1981) to be most accurate model available.

Water Hyacinth Modeling

Very little modeling of hyacinth growth has been reported in the literature. The most comprehensive work to date has been the model presented by Lorber et al. (1984). They proposed a basic physiological equation to describe hyacinth growth as follows:

$$\frac{dW}{dt} = [\ (P_g - R_m) \cdot E\ ] - D$$

where

W = dry weight, mass/area,

P_g = gross photosynthesis, mass/area,

R_m = maintenance respiration, mass/area,

E = conversion efficiency

D = detrital production, mass/area.

It was assumed that the hyacinths were in a vegetative stage, and phenological stages were not included.

Gross photosynthesis is primarily a function of incident solar radiation and temperature. Nitrogen and phosphate may become limiting factors. Density may be a limiting factor if the plant population becomes so dense

that shading occurs. The model was extensively validated with data from several sites in Florida.

Curry et al. (1987) attempted to add nutrient uptake to the work of Lorber et al. (1984). However, the model frequently underestimated the nitrogen uptake by 30% or more. Stewart et al. (1987) presented another model of nutrient uptake for water hyacinths. Their model used a Monod relationship for nitrogen and phosphorus uptake. The model was validated with data from five wastewater treatment systems. In general, the model was considered adequate for operational purposes. Nevertheless, additional information is needed to fully characterize nutrient uptake by water hyacinths.

Systems Modeling

Modeling may be described as compositional analysis (Smerage, 1982). There are two levels of systems modeling. The conceptual model is a qualitative statement of the components of a system, a generic description of those components, and a description of their interconecting relationships. Mathematical models form the second level and are comprised of mathematical statements which describe the system components and their interrelationships.

Behaviorial analysis is the determination and interpretation of the behaviorial properties of a system by analysis of a mathematical model. This may be carried out

by direct mathematical analysis for very simple models. However, computer simulation is required for the analysis of most real systems.

Simulations of agricultural and biological systems generally fall into one of two broad, and sometimes overlapping, catagories. Management oriented models generally involve the study of existing systems, the comparison of alternatives, or the design of new systems. Research oriented models are distinguished by objectives related to understanding the mechanism of behavior of a system.

Research models are unique in that they may be most valuable in their failure to predict the system behavior accurately, thereby disproving the hypothesis and pointing the direction for further research (France and Thornley, 1984). This statement is supported by Jones et al. (1987, p.16) in the statement, "a model cannot be validated, it can only be invalidated".

Jones et al. (1987, p.18) quotes Dent and Blackie (1979) in their excellent discussion on validation. In part, they proposed that a model is adequate if

a. the model is not different from the real existing system to a degree that will detract from the value of the model for the purposes for which it was designed.

b. That if the model is accepted as being adequate then the decisions made with its assistance will not be measureably less correct than those made without the benefit of the model.

Kloss (1982) reported on the use of a farm system model to determine the optimal layout for a biogas plant. Several options were investigated, and a number of general recommendations were presented.

A swine waste digestion system was modeled by Durand et al. (1987). The purpose of the model was to optimize the system for net energy production. Several recommendations were made based upon the model, including the use of psycrophilic digestion to reduce digester heat loss and the use of fuel cells for energy production in place of an engine-generator.

Walker (1984) presented a detailed description of the modeling process with special application to the modeling of a dairy farm. A particularly valuable point was brought out in the distinction between process integration and system integration in traditional energy analysis. Walker emphasized that focusing on a single process or group of processes may be counterproductive in shifting dependence from one material, such as fuel oil, to another, such as scarce and energy intensive alloys. It was emphasized that all components of the system, including management, must be examined with a view toward the system as a global entity.

METHODOLOGY

Model Development

General

A schematic diagram of the system is shown in Figure 5. Development of the system model is presented in six sections. First, the overall mass balances for hydraulic flow and volatile solids will be presented. Second, mass balances for the system products, algae, hyacinths, and biogas, will be presented. A discussion of the thermal energy balance will also be included in this section. Third, the costs and revenues associated with the system and their economic impact will be examined. The fourth, fifth, and sixth sections will describe in detail the simulation modeling of the bioconversion processes involved in the anaerobic digesters, the algal growth unit, and the aquatic plant unit, respectively.

System Flow and Volatile Solids Mass Balances

A diagram showing the hydraulic and volatile solids (VS) flows in the system is presented in Figure 6. The system hydraulic flow and VS content are defined by input data. Refer to Appendix A for nomenclature assigned to the system variables.

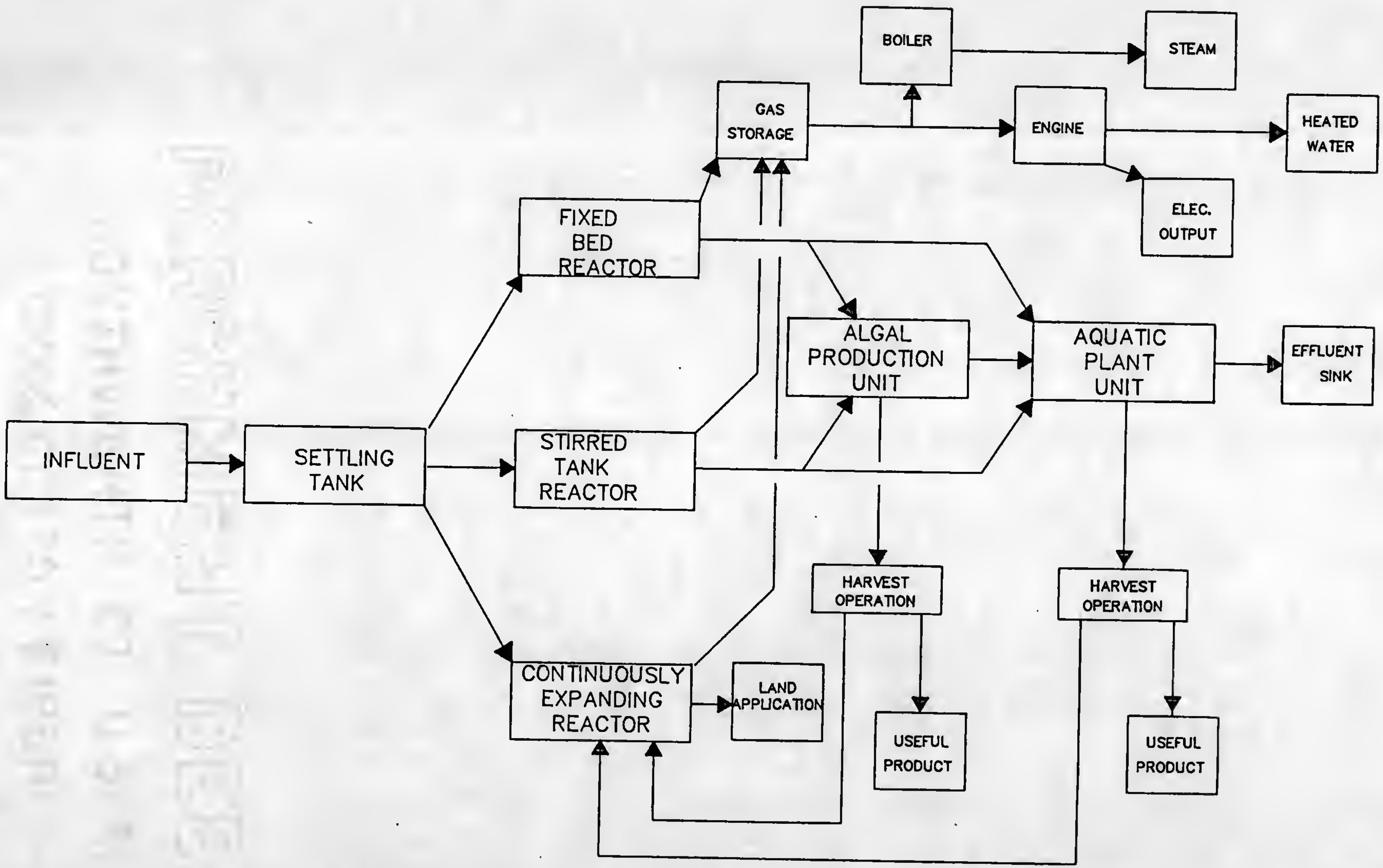

Figure 5. Block diagram of the integrated methane from biomass system.

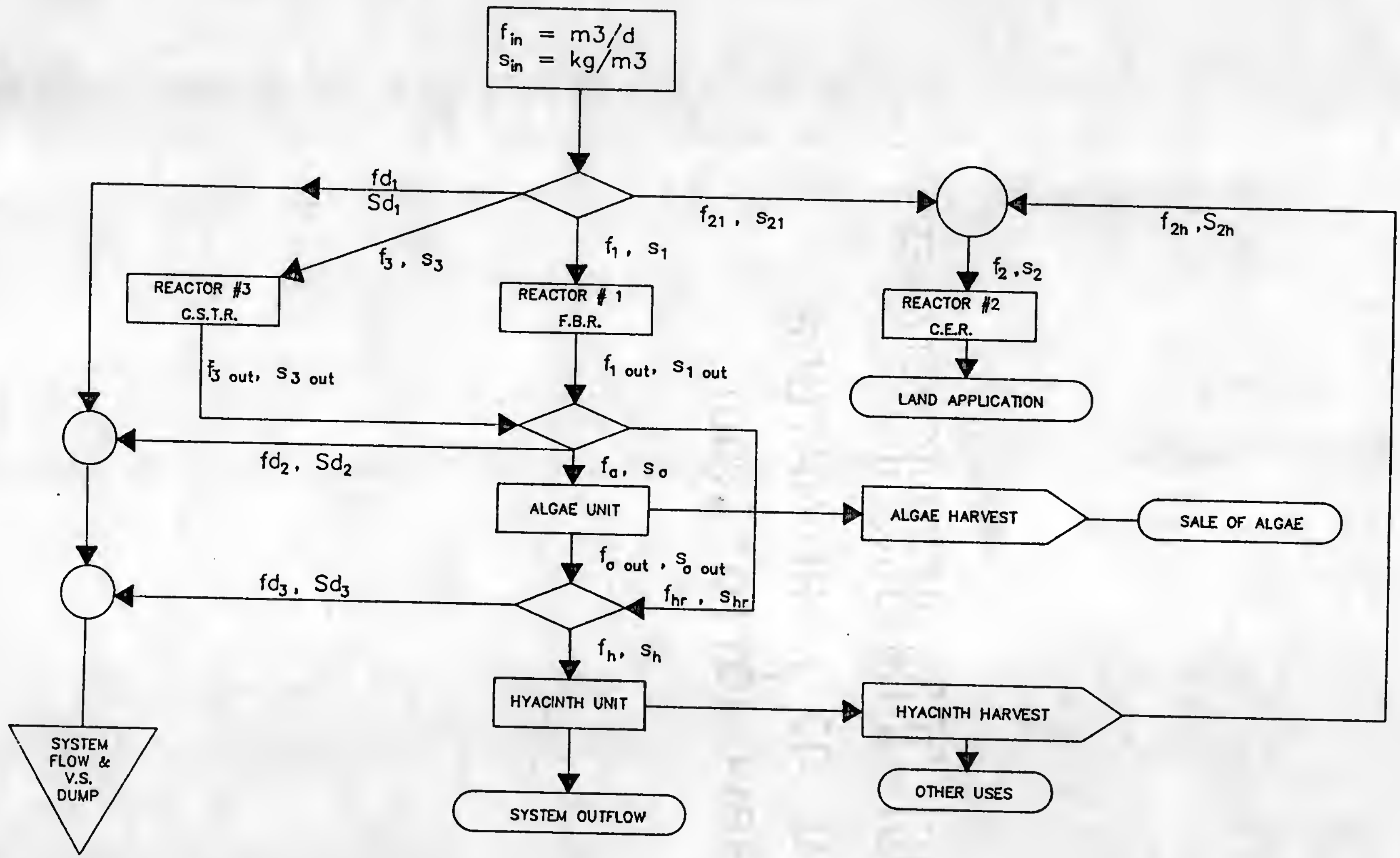

Figure 6. Mass flow diagram of the integrated methane from biomass system.

The first process to be modeled was the initial separation of the liquid and solid components of the influent stream and their distribution to the various reactors or to the system dump, or sink. The system dump represents a municipal sewer or other outfall which can accept theoretically unlimited hydraulic and VS flows with an economic penalty based upon both the hydraulic load and the VS load accepted. The mass balance for the initial flow distribution is as follows:

$$f_1 + f_{21} + f_3 + f_{d1} - f_{in} = 0$$

and

$$S_1 + S_{21} + S_3 + S_{d1} - S_{in} = 0$$

where f_{in} and S_{in} are externally specified and:

$$S_{in} = f_{in} \cdot s_{in} \qquad \text{(kg)}$$

$$f_1 = f_{in} \cdot f\%_1 \qquad (\mathrm{m^3/d})$$

$$f_{21} = f_{in} \cdot f\%_2 \qquad (\mathrm{m^3/d})$$

$$f_3 = f_{in} \cdot f\%_3 \qquad (\mathrm{m^3/d})$$

$$f_{d1} = f_{in} - f_1 - f_{21} - f_3 \qquad (\mathrm{m^3/d})$$

$$s_1 = \frac{S_{in} \cdot VS\%_1}{f_1} \qquad (\mathrm{kg/m^3})$$

$$S_{21} = S_{in} \cdot VS\%_2 \qquad \text{(kg)}$$

$$s_{21} = \frac{S_{21}}{f_{21}} \qquad (kg/m^3)$$

$$s_3 = \frac{S_{in} \cdot VS\%_3}{f_3} \qquad (kg/m^3)$$

$$S_{d1} = S_{in} - s_1 \cdot f_1 - S_{21} - s_3 \cdot f_3 \qquad (kg)$$

$$f_2 = f_{21} + f_{2h} \qquad (m^3/d)$$

$$s_2 = \frac{S_{21} + S_{2h}}{f_2} \qquad (kg/m^3)$$

where $f\%_n$ is the fraction of flow distributed to a unit and $VS\%_n$ is the fraction of VS distributed to a unit.

Flow from the reactors may be distributed to one of three locations. These are the algal production unit, the hyacinth growth unit, and the system dump. The fraction of flow sent to each is externally determined. The mass balance for this distribution is as follows:

$$f_1 + f_3 - f_{hr} - f_a - f_{d2} = 0$$

$$S_1 + S_3 - S_{hr} - S_a - S_{d2} = 0$$

where

$$f_a = f_1 \cdot f\%_{a1} + f_3 \cdot f\%_{a3} \qquad (m^3/d)$$

$$f_{hr} = f_1 \cdot f\%_{h1} + f_3 \cdot f\%_{h3} \qquad (m^3/d)$$

$$f_{d2} = f_1 + f_3 - f_a - f_{hr} \qquad (m^3/d)$$

$$s_a = \frac{f_1 \cdot f\%_{a1} \cdot s_{1_{out}} + f_3 \cdot f\%_{a3} \cdot s_{3_{out}}}{f_a} \quad (kg/m^3)$$

$$s_{hr} = \frac{f_1 \cdot f\%_{h1} \cdot s_{1_{out}} + f_3 \cdot f\%_{h3} \cdot s_{3_{out}}}{f_{hr}} \quad (kg/m^3)$$

$$S_{d2} = f_1 \cdot s_{1_{out}} + f_3 \cdot s_{3_{out}} - f_a \cdot s_a - f_{hr} \cdot s_{hr} \quad (kg)$$

The effluent from the algal production unit may be directed to the hyacinth pond or to the system dump. The fraction of effluent sent to the hyacinths is determined by input data. The mass balance for the distribution of effluent from the algal pond and for the hyacinth influent is given below.

$$f_{ha} = f_{a_{out}} \cdot f\%_{ha} \quad (m^3/d)$$

$$f_{d_3} = f_{a_{out}} - f_{ha} \quad (m^3/d)$$

$$f_h = f_{hr} + f_{ha} \quad (m^3/d)$$

$$s_h = (f_{hr} \cdot s_{hr} + f_{ha} \cdot s_{ha}) / f_h \quad (kg/m^3)$$

Effluent from the hyacinth growth unit is discharged from the system.

Product Mass and Energy Balances

Algae are harvested from the growth ponds at a specified frequency or when a predetermined density has been

reached. The amount of algae harvested is calculated on a dry weight basis according to the formula:

$$\text{algal harvest} = [\text{algal density before harvest } (g/m^3) - \text{algal density remaining}] \cdot \text{pond volume.}$$

The harvest is reported in tonnes (metric) dry weight. The algal mass balance for the system is:

$$\text{Algae @ } t=0 + \text{net algae produced} - \text{algae harvested} - \text{algae @ } t=\text{end} = 0.$$

Similarly, the harvesting of hyacinths occurs at either a specified frequecy or at a predetermined plant density. The harvest is calculated as:

$$\text{hyacinth harvest} = [\text{density before harvest } (g/m^2) - \text{density remaining}] \cdot \text{pond area.}$$

The harvest is reported in tonnes dry weight. The mass balance for the hyacinth system is as follows:

$$\text{hyacinths @ } t=0 + \text{net hyacinth production} - \text{hyacinths sold} - \text{hyacinths fed to the CER} - \text{hyacinths @ } t=\text{end} = 0$$

The biogas produced by the three reactors may be used to power an electrical generator or fire a boiler. It may also be stored for future use or, if the storage capacity, defined by input data, is insufficient, it may be flared off for safe disposal. Methane was assumed to have a heating value of 39.1 MJ/m^3.

Methane production from the digesters was summed as follows:

$$M_1 + M_2 + M_3 + M_s = M_t$$

$$G_1 + G_2 + G_3 + G_s = G_t$$

$$\frac{M_t}{G_t} \cdot 100 = Q_t$$

where G_n = biogas produced by reactor n, M_n = the methane produced from reactor n, t = total, s = stored, and Q_n is the percent methane content of the biogas. The mass balance was:

biogas produced - biogas consumed - biogas stored = 0.

When adequate biogas was available, the generator was assumed to produce its rated capacity for a 24 hour period. If insufficient biogas energy was available to run at full capacity then only the amount available was converted to electrical kilowatt-hours. Low grade heat in the form of hot water was available from the generator cooling system.

The hot water could have been used for digester heating or other in-plant uses.

Biogas in excess of that needed for electrical power generation was used to fire a boiler. If sufficient energy was available the boiler produced steam at its rated capacity. Otherwise it converted the biogas that remained after electrical generation into steam. Boiler steam could be used to heat the digesters, used in-plant or sold for off-site uses.

Any biogas remaining after both the generator and the boiler have operated at their full rated capacity was stored for future use. If insufficient storage capacity was available, the remaining gas was flared.

A thermal energy balance of the heated digesters was performed to determine the heat losses. These losses must be made up by heat from either co-generated hot water or from the boiler generated steam. If both of these sources were insufficient to maintain reactor temperatures, then heat energy was purchased from an outside source.

Two sources were taken into account during calculation of the reactor heat losses. These were the conductive losses through the walls of the reactor vessel and the convective losses due to warm effluent being replaced by cooler influent. In the first instance the losses were calculated using the vessel surface area and an effective "R" value for the reactor walls. If reactor temperature was

greater than the ambient temperature as defined by input data the heat losses were calculated by :

$$Q = \frac{\text{Area} \cdot (T_{reactor} - T_{ambient})}{R}$$

where Q is in units of Joules/sec.

The second source of heat loss is convective and was calculated by the equation:

$$Q = \text{mass flowrate} \cdot C \cdot (T_{reactor} - T_{influent})$$

where C is the specific heat of the substrate.

The influent temperature was assumed to be ambient. However, the system may be defined through the input data to include a heat exchanger. This reduces the temperature differential by an amount equal to the heat exchanger efficiency.

Costs, Revenues, and Economic Analysis

The costs associated with the system were of two types, capital costs associated with initial construction of the system, and operating costs.

Capital costs, other than land, were calculated in terms of 1976 dollars in order to properly account for inflation. These costs include digester construction, algal growth unit and aquatic plant unit construction, and a pumping station. Engineering, contingency and administrative costs were also included. These costs were adjusted to

current values by use of the Engineering News Record (ENR) Construction Cost Index, the current value of which is input data. The cost of land was derived from input data in current dollars.

Operating costs, including depreciation and interest costs, were calculated in current dollars. Unit costs for labor, electrical power, flocculating agents, and other operating expenses are input data.

Most of the equations used to estimate the capital costs were derived from information presented by Vernick and Walker (1981). Exceptions were the aquatic plant unit, the cost of which was derived from McMahon and Pereira (1984), and the algal unit. The methods used to estimate the capital cost of the major components are described below.

Vernick and Walker (1981) presented cost data in the form of graphs of construction cost (in 1976 dollars) vs. plant size for various waste treatment unit operations. Data points were taken from these curves and regressions performed to obtain the cost equations presented herein.

Capital costs associated with the construction of the anaerobic reactors were estimated by adapting information on the cost of anaerobic municipal sludge digesters. The equation:

$$\text{Cost} = \$6906 \cdot \text{Volume}^{0.494}$$

approximated ($r^2 = 0.97$) the data of Vernick and Walker (1981).

The cost of the algal production unit was calculated from the information presented by Johnson, et al. (1988) for algal raceways. When adjusted to 1976 dollars, the construction cost was approximately $89,400 per hectare. This fits well with the $80,000 - $100,000 per hectare reported by Dubinsky et al. (1978).

The growth ponds for the aquatic plant unit were estimated to cost $2.24 /m^3. This figure was based solely on the earthmoving, compaction, and embankment required to form the ponds.

The cost for construction of a pumping station was estimated using the equation:

$$\text{Cost} = \$486.5 \cdot \text{design flow } (m^3/d)^{0.58},$$

which matched the data of Vernick and Walker ($r^2 = 0.96$) and closely approximates the author's experience with sewage lift stations and water treatment plants.

Non-component capital costs, such as laboratory instrumentation, materials handling equipment, small buildings, and other miscellaneous capital expenditures were estimated to be 30% of the major component capital cost (Vernick and Walker, 1981).

Administrative costs were assumed to be 10% of the total material capital cost. Engineering and contingency costs were each defined as 15% of the material and administrative capital costs.

The base capital cost of the system was defined as the sum of the major component capital costs, non-component capital costs, and administrative, engineering, and contingency costs. This value was adjusted for inflation using the ENR construction cost index by the following formula:

$$\text{adj. capital cost} = \frac{\text{base cost} \cdot \text{new index}}{2475}$$

The cost of the land was added to the adjusted capital cost, resulting in the total capital cost of the system.

Depreciation was calculated on a straight line basis from the adjusted base cost. Simple interest was calculated on the total capital cost of the system.

Operating costs for the system were calculated in terms of current dollars. Calculation of the daily cost of operation was performed at the end of each day's simulation. These daily costs were summed over the simulation period to reflect a total operating cost for the simulated period.

The operating costs included the power cost of pumping between various units, mixing the CSTR and algal unit, and harvesting and processing hyacinths. Other costs included the purchase of energy to heat the digesters, chemicals to flocculate the algae during harvest, and charges associated with disposal of the effluent to the system dump, including surcharges for VS loading. The cost of labor, including both a fixed amount of labor to operate the system and a

variable amount dependent upon harvest size and frequency, was also calculated.

The power required for pumping was estimated to be 1.33 MJ/m^3 at 10m of head (Vernick and Walker, 1981). Reactor mixing requirements were assumed to be 14.9 J/m^3-sec. The daily power requirement for algal mixing, aeration, and harvest was defined as input data on a per unit volume basis. Power needed to harvest the water hyacinths varied directly with the quantity harvested and inversely with the average diameter of the chopped particles (Bagnall, personal communication).

The fixed labor cost for operating the system was based on the need for 3 full-time employees for plants up to 3785 m^3/d of design flow plus 1 additional employee for each additional 3785 m^3/d (1 mgd) of design flow. Variable labor costs were based on the number of tonnes of algae or hyacinths harvested and also on a fixed number of manhours per harvest for algae.

Revenues were calculated in terms of current dollars. As with the operating costs, revenues were calculated at the end of each day's simulation and summed to provide a total revenue for the period of the simulation. Revenues were assigned for the sale of algal and hyacinth biomass, electrical power and for steam from the boiler. Additional revenues were credited for the use of hot water from the co-generation system and for the removal of volatile solids,

which would otherwise incur a cost. When these credits were not to be considered in generating the economic summary, the respective values were set to zero in the input data.

A summary of the economic performance of the system was prepared at the end of each simulation. Net operating revenue was defined as total operating revenue plus total operating credits less total operating costs. Net revenue was defined as net operating revenue less interest and depreciation costs. The rate of return was calculated by the following equation:

$$\text{Rate of Return} = \frac{\text{Net Revenue}}{\text{Capital Cost}} \cdot 100$$

Tax implications were not considered in the economic analysis. Due to the complexity of the subject it was determined that tax implications were outside the scope of this work.

Anaerobic Digester Models

General process model

The anaerobic digestion process model was a generic model designed to represent a wide variety of reactor configurations. It was based on the "lumped parameter" method of Hill (1983a) and Hill et al.(1983). Hill (1983) used Monod kinetics for dynamic simulation. However, he simplified the characterization of wastes into only two

factors, the biodegradability factor (BO) and the acid factor (ACFACT).

Bolte (1985) adapted Hill's method to attached film reactors by incorporation of a bacterial retention coefficient (BRC). The BRC was the fraction of active biomass that was assumed could not be washed out of the digester.

Bolte worked with swine waste supernatant and assumed all biodegradeable organic matter was either soluble or so rapidly degradable as to be of negligible decay time.

The anaerobic digestion submodel created for use within this system model included a modification of the work of Bolte (1985) to include slowly degradable substrates and to provide better estimates of substrate effects on methane content of the biogas. It was designed to allow most reactor types to be represented with only minor changes, most of which could be made by changing parameters in a reactor specific procedure which called the process model as a subroutine. Integration was performed by a fourth order Runge-Kutta integration routine. The models of Hill (1983a) and Bolte (1985) were defined by only four state variables. These were the two bacterial populations and their respective substrates. The generic process model used in this study adds two additional state variables to account for varying degradation rates of complex organic matter. The state variables and their associated parameters are shown in Table 1.

Table 1. Selected variables used in the digestion model

m = acetogenic bacterial concentration, kg/m^3.

mc = methanogenic bacterial concentration, kg/m^3.

mu = acetogen growth rate, 1/d.

muc = methanogen growth rate, 1/d.

kd = acetogen death rate, 1/d.

kdc = methanogen death rate, 1/d.

lcom = lignified, or slowly degradable, complex organic matter, kg/m^3.

rdcom = readily degradable complex organic matter, kg/m^3.

lcombkd = breakdown rate of lcom, kg/m^3-d.

rdcombkd = breakdown rate of rdcom, kg/m^3-d.

SO = soluble organic material, kg/m^3.

VFA = volatile fatty acids, kg/m^3

BRC = bacterial retention coefficient

volk = volume of one compartment of a reactor, m^3

flow = volume of liquid added to the reactor, m^3/d

y = acetogenic yield coefficient, g org./g substrate

yc = methanogenic yield coefficient, g org./g substrate

yhac = acid yield coefficient, g VFA produced / g acetogenic bacteria produced

Digesters were divided into a number (k) of equally sized compartments which were considered to be completely mixed reactors. The six basic rate equations for k > 1 were as follows:

(1) Rate of change of acetogens:

$$\frac{dm[k]}{dt} = (mu - kd - (1-BRC)\frac{flow}{volk})m[k] + (1-BRC)\frac{flow}{volk}\ m[k-1]$$

(2) Rate of change of methanogens:

$$\frac{dmc}{dt} = (muc - kdc - (1-BRC)\frac{flow}{volk})mc[k] + (1-BRC)\frac{flow}{volk}\ mc[k-1]$$

(3) Rate of change of lignified COM:

$$\frac{dlcom}{dt} = (lcom[k-1] - lcom[k])\frac{flow}{volk} - lcombkd$$

(4) Rate of change of readily degradable COM:

$$\frac{drdcom}{dt} = (rdcom[k-1] - rdcom[k])\frac{flow}{volk} + lcombkd - rdcombkd$$

(5) Rate of change of soluble organics:

$$\frac{dSO}{dt} = rdcombkd + (SO[k-1] - SO[k])\frac{flow}{volk} + kd\cdot m[k]$$

$$+ kdc\cdot mc[k] - \frac{mu\cdot m[k]}{y}$$

(6) Rate of change of volatile fatty acids:

$$\frac{dVFA}{dt} = (VFA[k-1] - VFA[k])\frac{flow}{volk} + mu \cdot m[k] \cdot yhac - \frac{muc \cdot mc[k]}{yc}$$

The equations were modified slightly for the compartment k = 1. In this case the influent bacterial concentrations were assumed to be zero, and the substrate [k-1] concentration was replaced by the influent substrate concentration (ie. SOin).

The above equations represent the rate equations for continuous flow reactors. Non-continuous flow reactors such as the CER were modeled as a single homogeneous reactor with no effluent flow. The unloading of this type of reactor was treated as a discrete event and handled outside the digestion routine. The rate equations for the CER were as follows.

(1) Rate of change of acetogens:

$$\frac{dm}{dt} = (mu - kd) \cdot m$$

(2) Rate of change of methanogens:

$$\frac{dmc}{dt} = (muc - kdc) \cdot mc$$

(3) Rate of change of lignified COM:

$$\frac{dlcom}{dt} = (lcomin) \frac{flow}{volk} - lcombkd$$

(4) Rate of change of readily degradable COM:

$$\frac{drdcom}{dt} = (rdcomin)\,\frac{flow}{volk} + lcombkd - rdcombkd$$

(5) Rate of change of soluble organics:

$$\frac{dSO}{dt} = (SOin)\,\frac{flow}{volk} + kd\cdot m + kdc\cdot mc + rdcombkd - \frac{mu\cdot m}{y}$$

(6) Rate of change of volatile fatty acids:

$$\frac{dVFA}{dt} = (VFAin)\frac{flow}{volk} + mu\cdot m\cdot yhac - \frac{muc\cdot mc}{yc}$$

The bacterial growth and death rates were those proposed by Hashimoto (1980) and have been used extensively by others (Hill, 1983b; Dwyer, 1984) in validated models. Hashimoto proposed that a single maximum specific growth rate ($m\hat{u}$) be used which was common to all of the populations in the digester. He determined this value to be

$$m\hat{u} = 0.013\cdot(T) - 0.129$$

where T was the temperature between 20 and 60 degrees C.

The maximum specific bacterial death rate was proposed to be equal to the maximum growth rate (Hill et al., 1983). The $k\hat{d}$ had previously been taken as one tenth of $m\hat{u}$. However, it was shown that this was an insufficient removal mechanism in CER type reactors where there is no bacterial washout. The Monod coefficients ki, kic, ks, and ksc were taken directly from Hill et al. (1983). The following

equations were used to calculate specific growth and death rates.

$$mu = \frac{\hat{mu}}{1 + \frac{ks}{SO[k]} + \frac{VFA[k]}{ki}}$$

$$muc = \frac{\hat{mu}c}{1 + \frac{ksc}{VFA[k]} + \frac{VFA[k]}{kic}}$$

$$kd = \frac{k\hat{d}}{1 + \frac{ki}{VFA[k]}}$$

$$kdc = \frac{k\hat{d}c}{1 + \frac{kic}{VFA[k]}}$$

where mu = acetogenic growth rate (d^{-1}), muc = methanogenic growth rate (d^{-1}), kd = acetogenic death rate (d^{-1}), and kdc = methanogenic death rate (d^{-1}).

The degradation rates for rdcom and lcom were based on simple first order kinetics. Rdcom was defined as that material which had an average half-life of two days before breakdown into soluble organic material. The two day half-life was determined during calibration and sensitivity testing using reactors operated at retention times which varied from 6 days to 9 hours.

Similarly, lcom was defined to have a half-life of 20 days and break down into rdcom. This represented material

which was protected from rapid hydrolytic attack by the lignocellulosic structures frequently found in crop residues and other sources of biomass. Calibration of the lcom breakdown rate was accomplished using data from Chynoweth et al. (1984).

Yield coefficients for the model were obtained from a variety of sources. The acetogenic and methanogenic yield coefficients, y = 0.1 grams of acetogenic organisms / gram substrate, and yc = 0.0315 grams of methanogenic organisms / gram substrate, were obtained from Hill et al. (1983).

Yield of acetate from soluble organics metabolized during acetogenesis was derived from the stoichiometry of Dwyer (1984). This work was done on a model using five bacterial populations instead of two. Therefore, contributions from the degradation of propionate and butyrate, and from the homoacetogenic conversion of CO_2 were added to obtain the proper value. Because neither pH or carbonate balances nor hydrogen-using methanogens were considered in this model, it was assumed that approximately 11% of the available CO_2 from acetogenesis was used in homoacetogenesis. This was supported by sensitivity analysis during model calibration. A value of yhac = 6.64 grams VFA / gram acetogens produced was used in the model.

The volumetric yield of CO_2 from acetogenesis was taken directly from Hill and Barth (1977). The volumetric yields of methane and CO_2 from methanogenesis were modifications of

the Hill and Barth parameters. To adjust the volumetric yields of Hill and Barth to account for a higher proportion of hydrogen utilizing methanogenic bacteria when non-carbohydrate substrates were used, a factor, CH4RAT, was defined. This was defined as 1.0 for pure carbohydrate substrates but varied down to 0.69 for swine waste. It may be roughly thought of as having represented the conversion of CO_2 to acetate and the direct reduction of CO_2 to methane.

The volumetric yield parameters were as follows:

$yvCO_2 = 2.35$,

$yvCH_4 = 15.86 / CH4RAT$,

$ycvCO_2 = 9.32 - [15.86 (1-CH4RAT)]$,

where $yvCO_2$ = L CO_2 / gram acetogens produced, $yvCH_4$ = L CH_4 / gram methanogens produced, and $ycvCO_2$ = L CO_2 / gram methanogens produced.

Volatile solids reduction was assumed to be through conversion to CO_2 or methane. The following equation represents this destruction:

VS destroyed = 0.782 (VFA used) + 0.284 (SO used).

The conversion factors were from two sources. The conversion factor for the destruction of VFA by methanogens, 0.782, was taken directly from Bolte (1985). The acetogenic destruction of SO was calculated from the work of Dwyer (1984).

Substrate composition has a marked effect on digester operation. The values of BO and ACFACT used to describe

beef, dairy, swine, and poultry waste were those of Hill (1983a). Bolte (1985) described screened swine waste (ACFACT = 0.10 and BO = 0.95) and protein-carbohydrate waste (ACFACT = 0.001 and BO = 1.00).

Inherent in the use of only two "lumped" parameters to describe a substrate was the assumption that the substrate was instantly available for use by the bacterial population. This assumption may have been valid for relatively volatile wastes digested at long retention times. In such a case the time required for hydrolysis is short compared to the retention time.

Where these conditions were not met, as in a fixed bed reactor operating at a short retention time, or when digesting slowly degradable biomass, it became necessary to devise a more accurate representation.

To satisfy this need to represent the breakdown of complex materials, two additional substrate parameters were defined. These were the readily degradable fraction of the biodegradable solids (rdf) and the lignified, or slowly degradable, fraction of the biodegradable solids (lf).

The readily degradable fraction of the waste was defined to have an average half-life of 2 days. This meant that there was little effect on reactor operation at retention times in excess of six to eight days, where most CSTR's operate. However, it had a very large effect on FBR's where retention times were much shorter and potential

substrate material could wash out of the reactor before it could be digested.

Calibration of the rdcom parameter for swine waste was accomplished using the data of Nordstedt and Thomas (1985b). Swine waste was determined to have a rdf = 0.69. The protein - carbohydrate waste was found to have a rdf = 0.8. The lignified component of these wastes was considered negligible. Beef and poultry wastes were not calibrated for rdf because low retention time data was unavailable. The rdf for beef was assumed to be 0.7 in this model based on similarity with swine waste. However, the lignified fraction could be significant in some cases where large amounts of bedding were incorporated. Poultry wastes were assigned a value for rdf = 0.5 as an estimate of their degradability, but this has not been tested at the short retention times needed for validation.

Continuous stirred tank reactor model

This is the conventional continuous flow anaerobic digester. All state variables were kept in their intensive (kg/m^3) form.

The model defines the bacterial retention coefficient as zero. The number of compartments to be simulated was defined as 1 since this was a theoretically homogeneous reactor. The waste type and influent concentration, flow, reactor volume and temperature were determined from input

data. These values, and those of the state variables at the end of the last day's simulation, were passed to the general process model. The returned state variables were stored for use in the next day's simulation.

Fixed bed reactor model

This model was identical to the CSTR in that it was a continuous flow reactor model using intensive variables. Operation was similar to the CSTR except for two parameters.

The FBR model defined an FBR as five compartments in series, each one-fifth of the total reactor void volume, so as to create a psuedo plug flow effect. The bacterial retention coefficient was BRC = 0.995. This resulted in an active bacterial retention time of 50 days at a hydraulic retention time of 6 hours. This was a reasonable value for a well developed biofilm, as evidenced by the stability exhibited by the number of successful reactors operated at short hydraulic retention times.

Continuously expanding reactor model

The continuously expanding reactor was treated differently from the types previously discussed because it was not a continuous flow reactor. The effluent term of the mass balance equation was zero except during discrete emptying events. The influent term, however, was semi-continuous. As a result the volume varied with time.

To correctly represent these differences the CER model stored state variables in their extensive form. Upon execution, the model calculated the new volume of the CER and converted the state variables from the extensive form to the intensive form used by the generic digestion model.

After the digestion routine had simulated the biological processes for the day, the returned state variables were multiplied by the volume to convert them to extensive form. They were then passed back to the global system model to be stored until the next execution of the CER simulation procedure.

Because there was no effluent, the bacterial retention coefficient was meaningless. The CER was assumed to be homogeneous and was simulated with a single compartment.

Because of the limited data available on CER operations and on hyacinth digestion, water hyacinth characteristics were specifically calibrated for the CER model. Biodegradability was taken to be 0.66 based on the work of Chynoweth et al. (1984). The readily degradable and lignified fractions were calibrated in accordance with the work of Nordstedt (1988). The parameters were set to rdf = 0.5, lf = 0.4, ch4rat = 0.7, and the acid factor = 0.05.

The discrete emptying event was triggered by the CER reaching the design maximum volume. First the state variables were converted to the intensive form. The CER volume was then reset to a predetermined "seed" volume to

begin the next cycle. Finally the state variables were returned to their extensive form. Effluent distribution for land application was also performed at this time.

Algal Growth Model

In general, growth and nutrient uptake kinetics of microalgae have been modeled using some form of the Monod relationship. Temperature has historically been used to modify the growth rate by using a linear or exponential temperature factor (Bolte et al., 1986). The algal growth model used here was based on the work of Hill and Lincoln (1981).

The model developed by Hill and Lincoln (1981) used five inputs to determine specific growth rate. These were temperature, CO_2, ammonia, orthophosphate, and light. The temperature effect was modeled as a standard Arrhenius function. The other nutrients, and light, follow the Monod form:

$$mu_{sub} = \frac{\text{Substrate}}{k_{s_{sub}} \cdot \text{Substrate}}$$

The most limiting substrate was used to determine the overall growth rate for the model at each iteration. A death rate, k_d was included to account for senescence and predation. The following equation was used to calculate the net algal growth rate.

$$mu_{algae} = (\hat{mu}_{algae} \cdot mu_{limit} - k_d) \cdot 1.05^{(T-25)}$$

where mu_{limit} was the most limiting mu_{sub}.

In experiments using wastewater from a swine lagoon it was determined that CO_2, ammonia and phosphate were present in excess (Hill and Lincoln, 1981). Since the primary feed for the algal unit was digester effluent, this model assumed that these nutrients were present in excess. Tracking the above nutrients through the system was left for further research. Only the radiation component was considered limiting in this model.

The daily radiation was considered to be spread over 13 hours in a sinusoidal pattern from 7 AM to 8 PM. K_s was determined to be 2.60 MJ/m_2-hour. This was equivalent to the value (1.037 Langleys/minute) used by Hill and Lincoln. The maximum specific growth rate was 3.0 d^{-1} as determined by Hill and Lincoln. In calibrating the model to the published data, k_d was set to 0.25 d^{-1}.

The rate equation d alg/dt = $mu_{algae} \cdot X$ was integrated hourly by a 4th order Runge-Kutta integrator to yield the algal concentration.

Algal yield in g/m^2 was determined by the effective depth of the algal channel. This was the depth to which adequate light could penetrate to sustain growth. Under normal conditions it varied between about 0.2 and 0.4 meters but could be substantially reduced by high turbidity.

Water Hyacinth Growth Model

The growth model for the water hyacinths (Eichornia crassipes) was taken from the work of Lorber et al. (1984). The basic physiological equation describing the growth of water hyacinths was:

$$dW/dT = (P_g - R_m) \cdot E - D$$

where W = dry weight, g/m^2

P_g = gross photosynthesis, g/m^2-day

R_m = maintenance respiration, g/m^2-day

E = conversion efficiency, dimensionless

D = detrital production, g/m^2-day

It was assumed that the plants remained in the vegetative stage, therefore, phenological growth stages were not modeled. It was felt that this assumption was valid due to the frequent harvesting which takes place in a biomass production system.

Gross photosynthesis (P_g) was a function of the amount of solar radiation intercepted by the plants, the temperature, and the nutrient levels in the ponds. It was described by the following equation:

$$P_{gmax} \cdot f(dens) \cdot f(T) \cdot f(P) \cdot f(N)$$

where

$$P_{gmax} = 22.318 + 0.102 \cdot S \qquad S > 100$$

$$= 0.32 \cdot S \qquad S \leq 100$$

and S = incident solar radiation in Langleys / day.

The density function accounted for the inability of the canopy to intercept all of the incident light at low densities. The function was based on work by Debusk et al. (1981) and was given by the equation:

$$f(dens) = W / 1000 \qquad W < 1000\ g/m^2$$
$$= 1.0 \qquad W \geq 1000\ g/m^2$$

The temperature function used by Lorber et al. (1984) and reported by Mitsch (1975) was modified slightly to promote model stability at low temperatures. The modified function was given as:

$$f(T) = 1.0 - 0.0038 \cdot (T-29)^2 \qquad T \geq 15\ C$$
$$= 0.255 \qquad T < 15\ C$$

In this model, phosphorus and nitrogen were assumed to be present in excess of the minimum concentrations. Therefore, f(P) and f(N) were defined as 1.0 in this model. In fact, the more likely scenario would be inhibition due to ammonia toxicity. However, this toxic effect is not well documented. Many other factors should also be considered in the removal of nutrients from ponds, several about which little is quantitatively known. Because of this, nutrient removal by the hyacinths was left for future modeling efforts.

The maintenance respiration requirement was extracted from gross photosynthate (P_g - R_m) prior to conversion to plant material. The respiration function was a linear function of the existing crop and was given as:

$$R_m = R_o \cdot W$$

where R_o was the respiration coefficient. In this model, R_o was set to the value 0.01 recommended by Lorber et al. (1984). This value was reasonable and in line with those of soybeans and other crops.

Because only the vegetative growth stages were modeled, the conversion efficiency (E) was considered a constant. Lorber et al. reported a conversion efficiency of 0.83, which was comparable to the 0.73 reported for soybean growth (Wilkerson et al.,1983).

The formation of detrital material due to overcrowding was estimated to occur at densities in excess of 2400 g/m^2. At a density of 2600 g/m^2 detrital formation was expected to equal the photosynthetic growth. The following equation governs detrital production due to overcrowding:

$$D = dW \cdot [1 - ((2600 - W) / 200)] \qquad W > 2400$$
$$= 0.0 \qquad W \leq 2400$$

It should be noted that production of detritus due to poor growth conditions was reflected in the term (P_g - R_m).

Model Validation

Model validation is the process by which the suitability of the model is evaluated in light of the purposes for which it was designed. As was mentioned at the end of the literature review, a model can never be validated, only

invalidated. The difference between calibration and validation is that during calibration the parameters of the model are adjusted to make the model fit the data. In the validation process the data used is independent of that used to calibrate the model and no adjustments are made to the model parameters. Only the model inputs which characterize the nature of the experiment producing the real data are manipulated. The output data from the simulation is then compared to the real data to obtain an indication of the validity of the model.

Continuous Stirred Tank Reactor Model

The CSTR model was validated by comparing the simulated methane production with that reported in the literature. Nineteen studies were evaluated. Swine waste was used as the substrate in 10 studies, 3 used beef waste, 2 used dairy waste, and 4 used the waste from poultry layer operations.

The type of waste, operating temperature, hydraulic retention time, influent volatile solids concentration and reference are listed in Table 2. The scatter diagram of predicted vs. actual volumetric methane production for the CSTR studies is presented in Figure 7. Linear regression was performed on this data, and the 95% confidence limits are shown by the envelope bounded by smooth curves on either side of the diagonal line representing perfect agreement.

Table 2. Sources for CSTR model validation data

Waste type	Temp. (C)	HRT (days)	VS (g/L)	CH_4 (L/L-d)	Reference
Swine	35	15	50.4	1.22	Hashimoto(1983)
Swine	55	15	50.4	1.45	Hashimoto(1983)
Swine	55	10	50.4	1.80	Hashimoto(1983)
Swine	22.5	40	36	.29	Stevens & Schulte(1979)
Swine	35	15	39.2	1.07	Fischer et al.(1975)
Swine	35	15	60	1.36	Fischer et al.(1975)
Swine	35	15	46.8	1.17	Fischer et al.(1975)
Swine	35	15	43.4	1.08	Fischer et al.(1975)
Swine	35	18	36.0	.71	Lapp et al.(1975)
Swine	35	30	31.4	.49	Kroeker et al.(1975)
Layer	35	44	69.1	.58	Converse et al.(1977)
Layer	35	31	59.5	.74	Converse et al.(1977)
Layer	35	42	81.9	.77	Converse et al.(1977)
Layer	35	52.5	72.5	.67	Converse et al.(1977)
Beef	55	12	62.5	1.59	Hashimoto et al.(1979)
Beef	55	7	82.6	3.57	Hashimoto et al.(1979)
Beef	35	20	47.5	.69	Burford et al.(1977)
Dairy	35	12	76.8	.77	Coppinger et al.(1978)
Dairy	35	15	64.7	.67	Converse et al.(1977a)

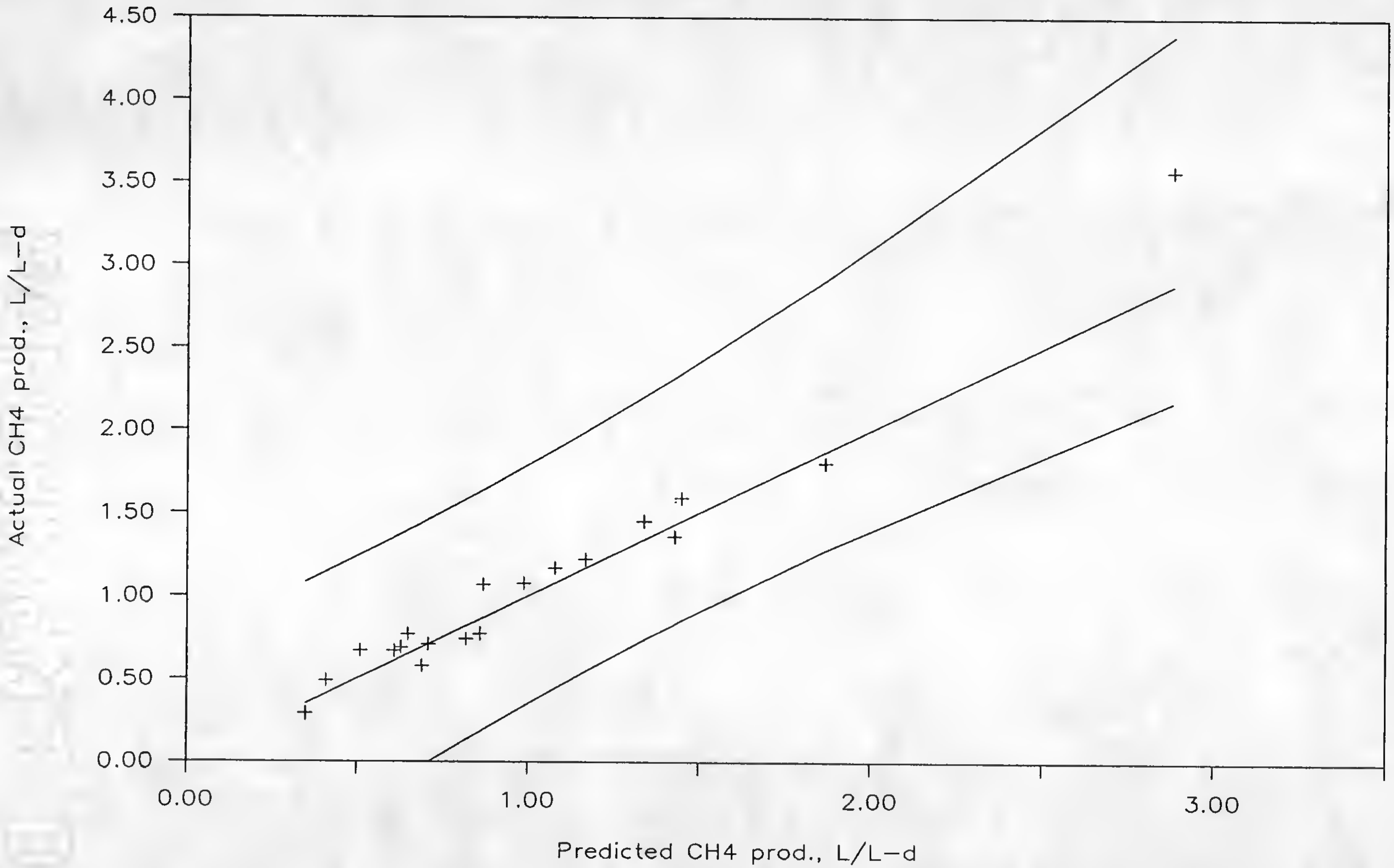

Figure 7. Predicted vs. actual volumetric methane production with 95% confidence limits for the CSTR model.

Fixed Bed Reactor Model

The fixed bed reactor model was validated by conducting simulations of 23 fixed bed reactor operations reported in the literature. Temperatures ranged from 24 to 55 C and hydraulic retention times from 1 to 19 days. Influent volatile solids concentrations varied from 1.5 to 25 kg/m^3. Substrates included both swine waste and food processing waste.

The type of waste, operating temperature, hydraulic retention time, influent volatile solids concentration and reference are listed in Table 3. A scatter diagram of predicted vs. actual volumetric methane production for the FBR studies is presented in Figure 8. Linear regression was performed on this data, and the 95% confidence limits are shown.

Continuously Expanding Reactor Model

Validation of the system's CER model was limited by lack of available data. The work of Hill et al. (1983) and Hill et al. (1985) provided many of the parameters used in the general digestion model. The only other CER studies available were those of Young (1979) and Nordstedt (1988).

The data of Nordstedt (1988) was used with the CER model to characterize the breakdown characteristics of the water hyacinth biomass. Therefore, this work was more

Table 3. Sources for FBR model validation data

Waste type	Temp. (C)	HRT (days)	VS (g/L)	CH_4 (L/L-d)	Reference
Swine	35	5	9.8	.84	Bolte et al. (1985)
Swine	35	2.5	9.3	1.38	Bolte et al. (1985)
Swine	35	1.5	10.3	2.0	Bolte et al. (1985)
Swine	55	2.5	10.4	1.84	Bolte et al. (1985)
Swine	55	1.5	9.6	2.68	Bolte et al. (1985)
Swine	55	1.0	10.0	3.29	Bolte et al. (1985)
Swine	24	1.0	4.0	.81	Brumm (1980)
Swine	24	2.0	3.8	.66	Brumm (1980)
Swine	24	3.0	3.3	.51	Brumm (1980)
Swine	24	6.0	3.9	.44	Brumm (1980)
Swine	35	1.0	4.0	.98	Hasheider & Sievers (1983)
Swine	35	3.0	1.5	.16	Hasheider & Sievers (1983)
Swine	35	3.0	3.0	.35	Hasheider & Sievers (1983)
Swine	35	3.0	6.0	.66	Hasheider & Sievers (1983)
Swine	35	3.0	12.0	1.25	Hasheider & Sievers (1983)
Swine	35	1.7	17.3	2.6	Kennedy & van den Berg(1982)
Swine	35	2.7	17.8	2.2	Kennedy & van den Berg(1982)
Swine	35	1.0	18.7	3.8	Kennedy & van den Berg(1982)
Swine	35	3.7	17.0	2.1	Kennedy & van den Berg(1982)
Bean	35	19	25	.35	Stevens & van den Berg(1981)
Bean	35	8.8	25	.86	Stevens & van den Berg(1981)
Bean	35	5.5	25	1.25	Stevens & van den Berg(1981)
Whey	32	2.0	5.6	.52	Thomas (1984)

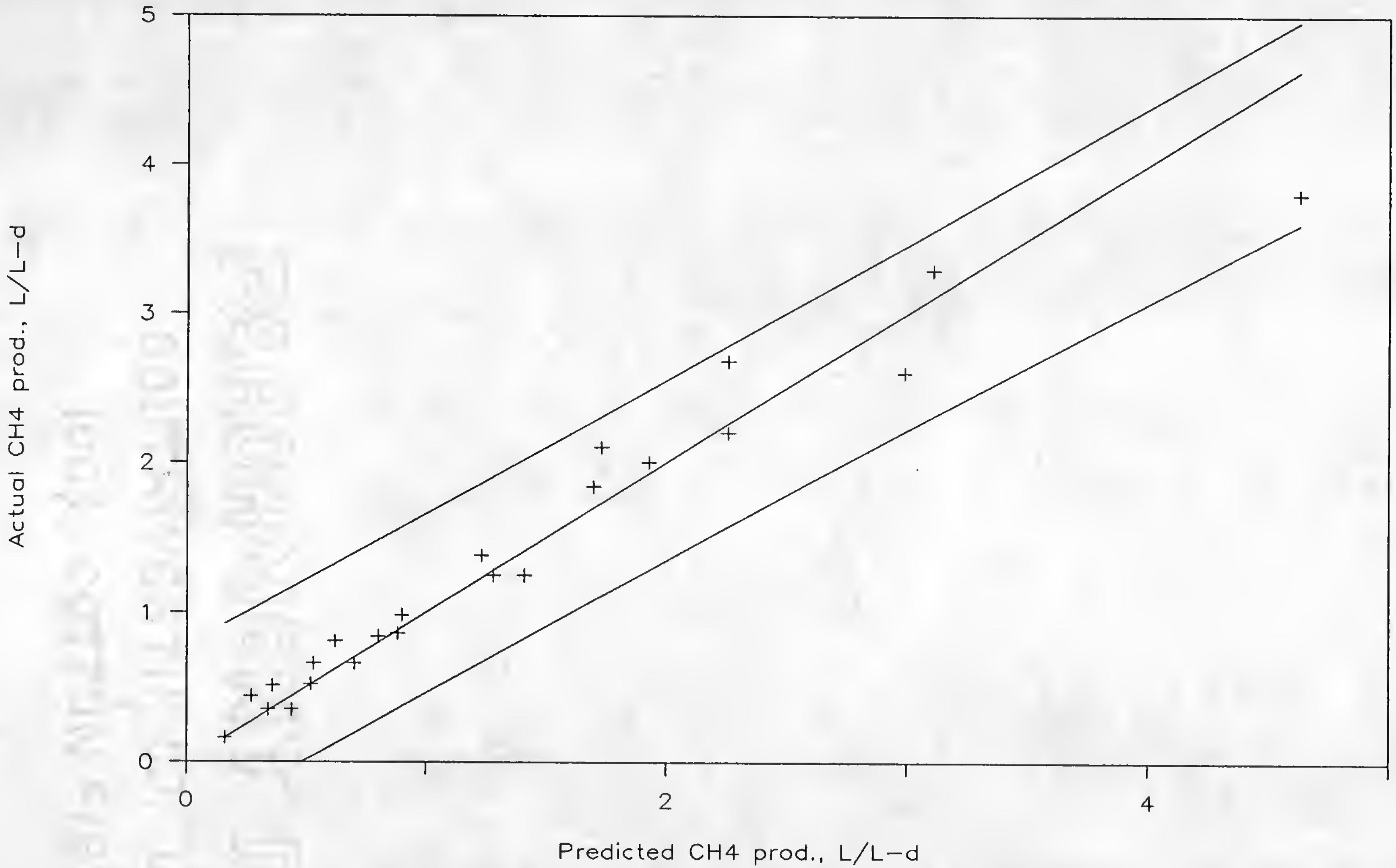

Figure 8. Predicted vs. actual volumetric methane production with 95% confidence limits for the FBR model.

properly characterized as calibration than validation. Actual data and predicted digester performance over time are shown in Figure 9. A scatter diagram of predicted vs. actual volumetric methane production for this study is presented in Figure 10. Linear regression was performed, and the 95% confidence limits are shown.

The work of Young (1979), conducted at 21 and 35 C in 1.4 m^3 digesters, was also simulated. Predicted long term yields of methane correlated very well with the actual data. Transient behavior was less accurate, particularly at a temperature of 21 C.

In Young's work, feeding was completely stopped at day 35 and not resumed until day 42, at which time a shock load of 7 times the normal daily load was fed to the CER. The model overestimated the effect of this transient. This was especially pronounced in the 35 C Trial (Figure 11). In addition, the model failed to predict a plateau in the methane production after the first 20 days in the 21 C trial (Figure 12).

Without additional data, it cannot be said that the CER model has been validated. However, poor transient response for a single study cannot be considered to have invalidated the model. As more work is done with the CER concept, additional simulations should be conducted to increase the confidence in the CER model.

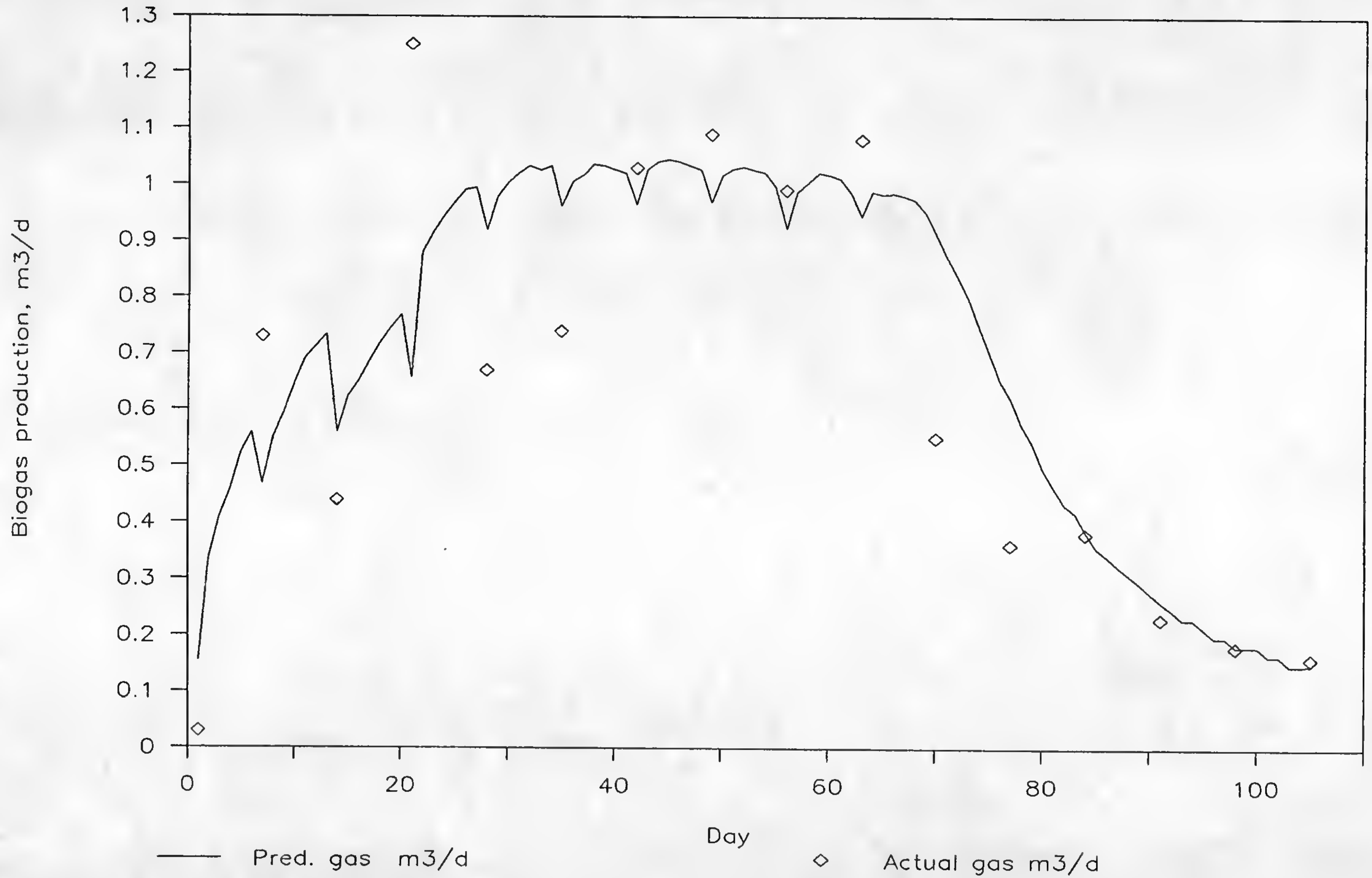

Figure 9. Simulation of CER hyacinth digester. Data from Nordstedt (1988).

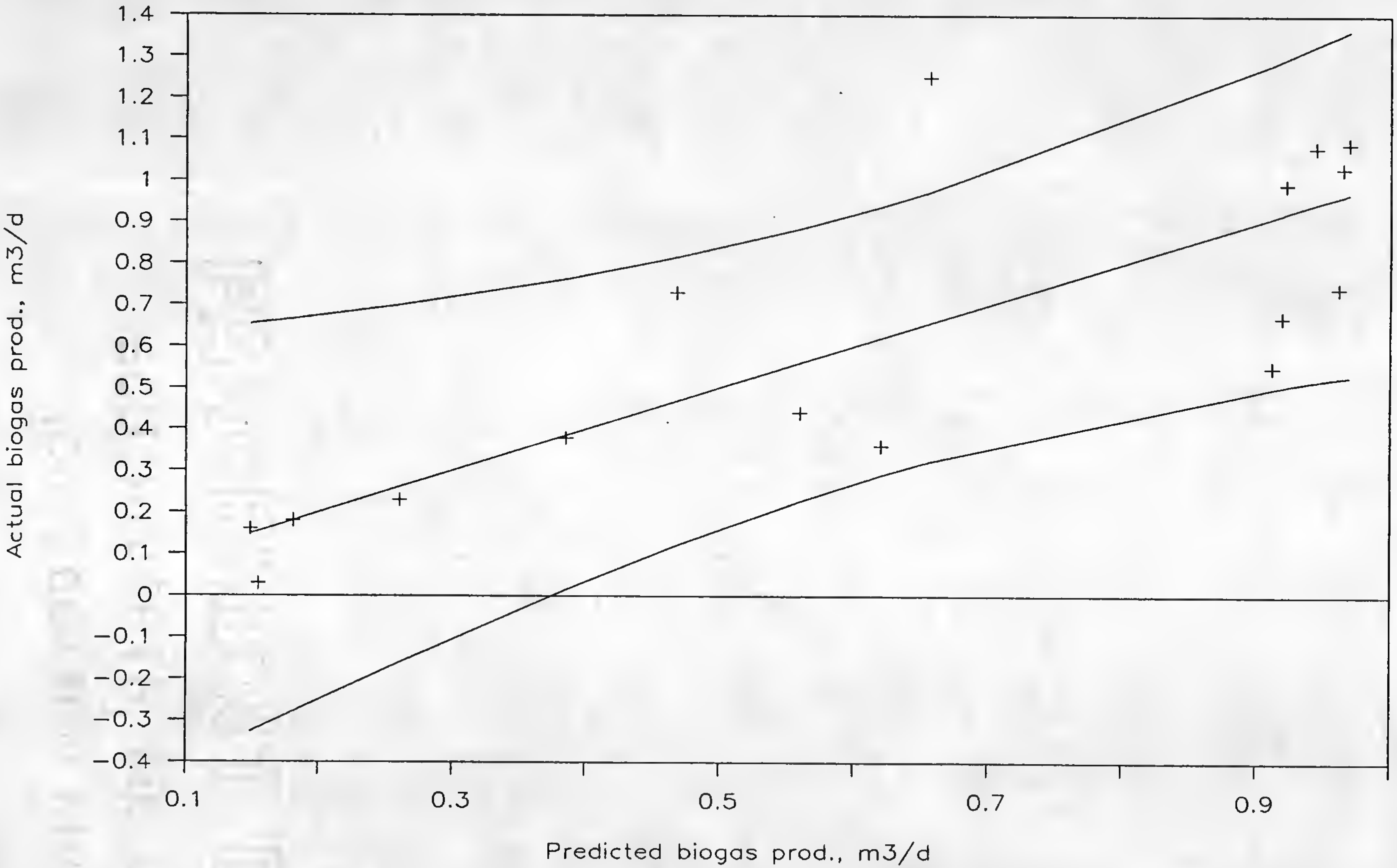

Figure 10. Predicted vs. actual biogas production with 95% confidence limits for the CER model. Data from Nordstedt (1988).

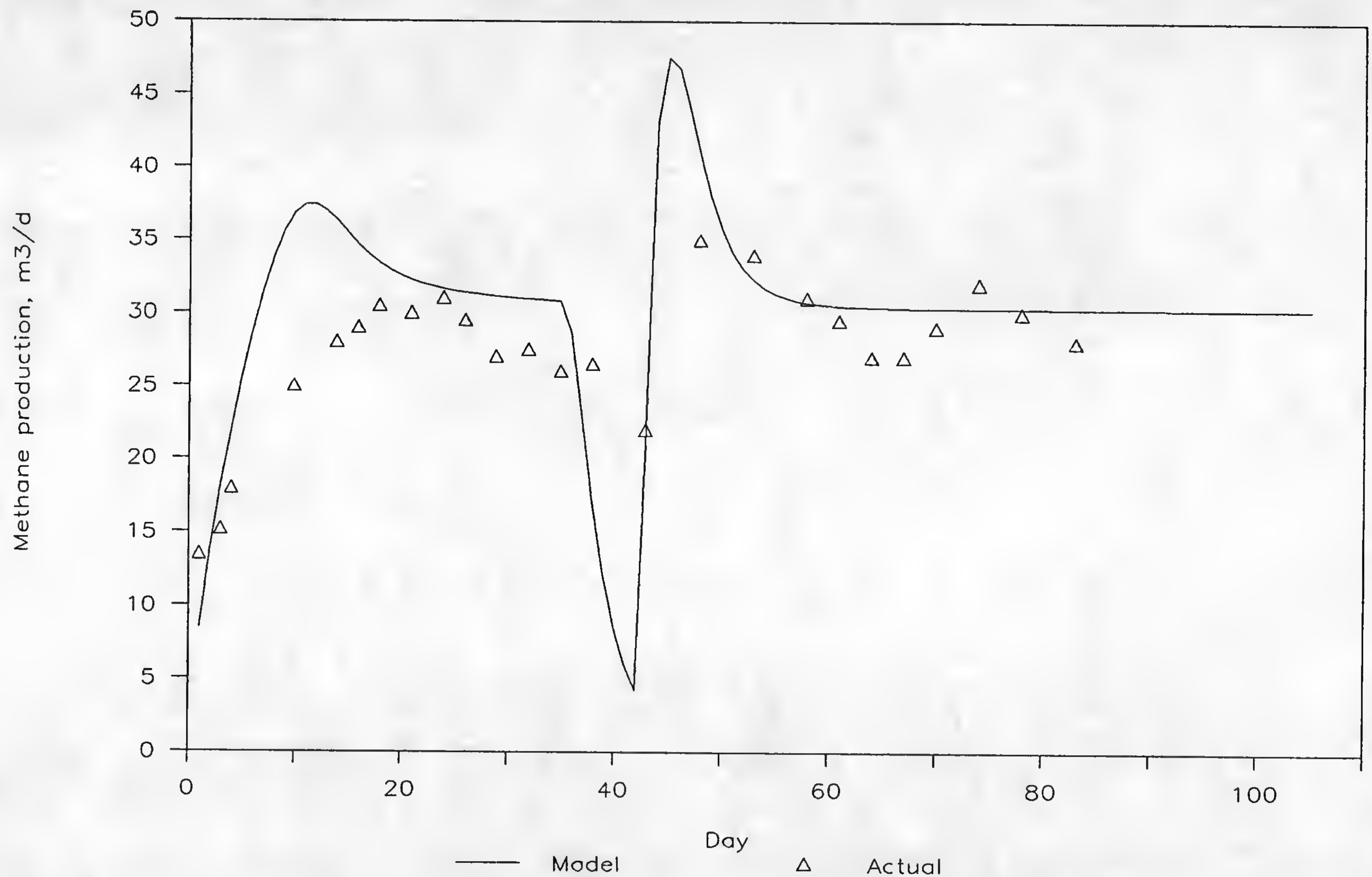

Figure 11. Simulation of CER using data from Young (1979) at 35 C.

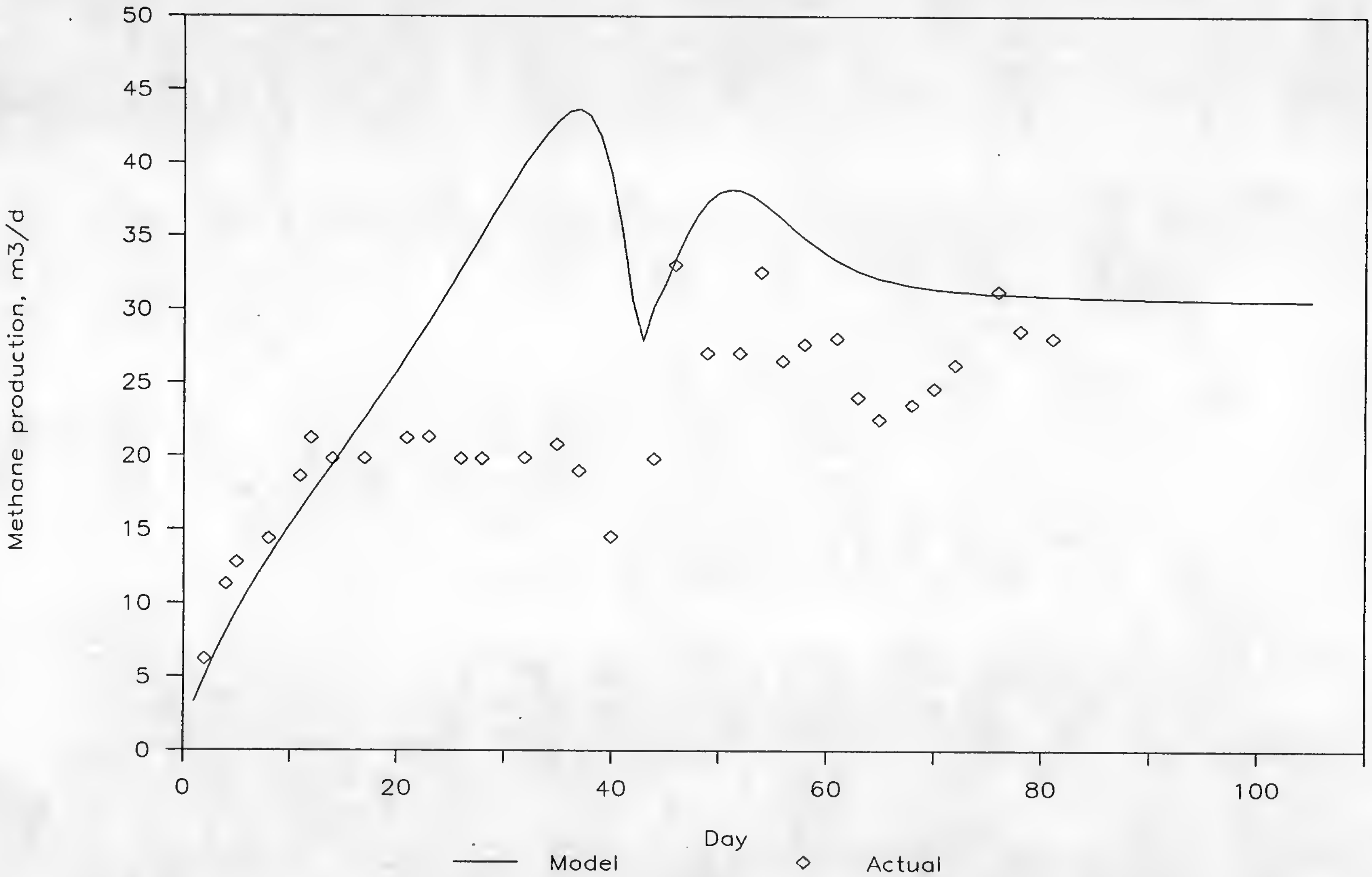

Figure 12. Simulation of CER using data from Young (1979) at 21 C.

Algal Growth Model

The algal growth model was based upon the parameters of Hill and Lincoln (1981) for determining growth and death rates. The model was calibrated using their published data. Thirteen hours of daylight was assumed with a half sinusoidal distribution over that period.

Validation of the model was performed by comparison of reported algal yields from thirteen studies conducted between 1967 and 1988 (Table 4).

Due to the lack of reported information about solar insolation levels and water temperatures in many references, most simulations were run using climatological data (Lunde, 1980; Landsberg, 1981; and Goldman, 1979b).

Algal yields for the various studies as well as the predicted value for each latitude are shown in Figure 13. Latitude was chosen for the X-axis to separate the experiments by location. It should be noted that most of these experiments were for less than a full year and were used only for model validation. They should not be taken as an indication of annual production. A scatter diagram of predicted vs. actual yields is given in Figure 14. The test for lack of fit was not significant at the 95% level.

A simulation of one year's production at Haifa, Israel was compared with actual data from Moraine et al. (1979) and is shown in Figure 15. The shift between the summer growth

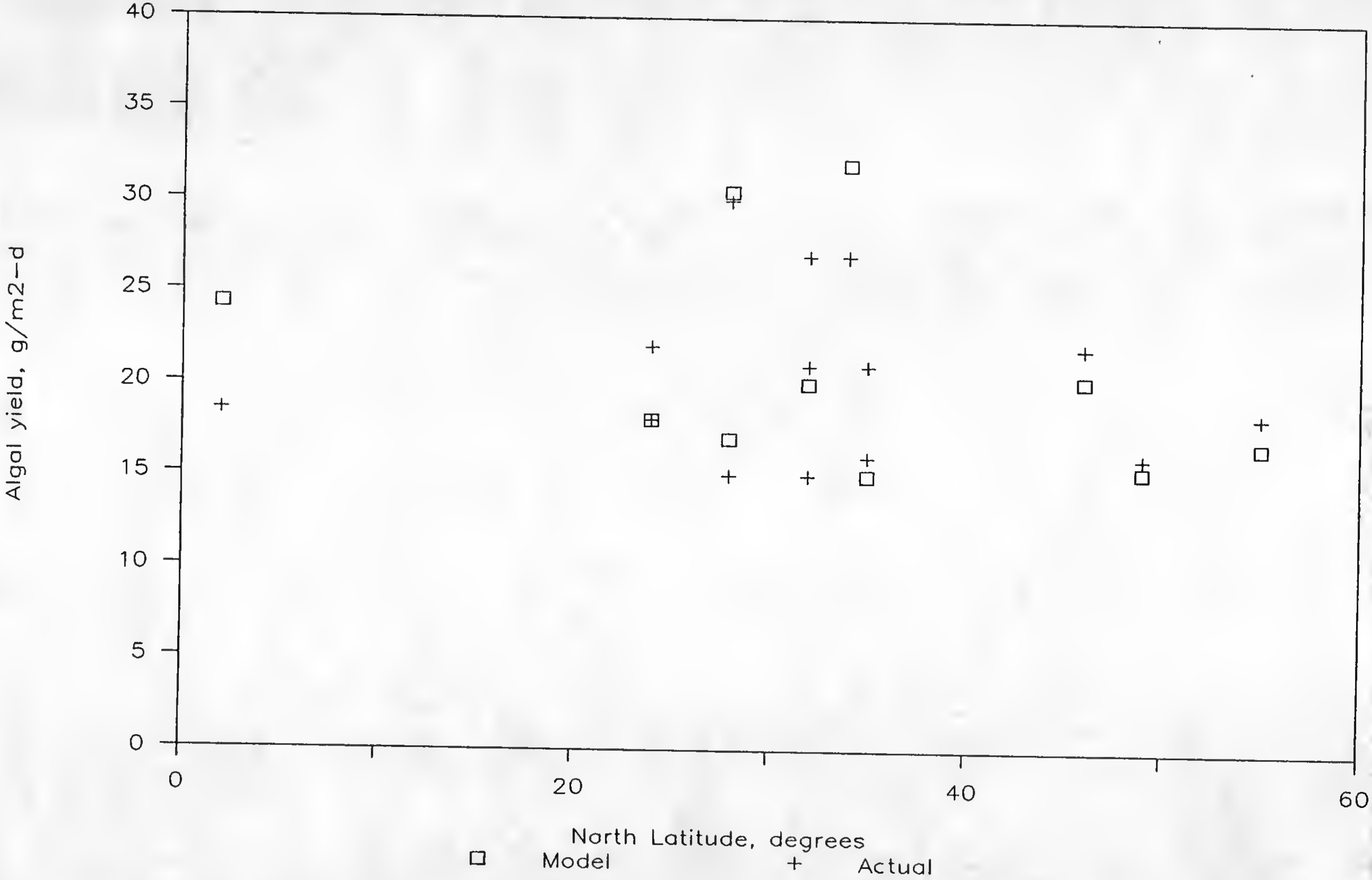

Figure 13. Actual algal yields and model predictions vs. latitude. Data was taken from studies referenced in Table 4.

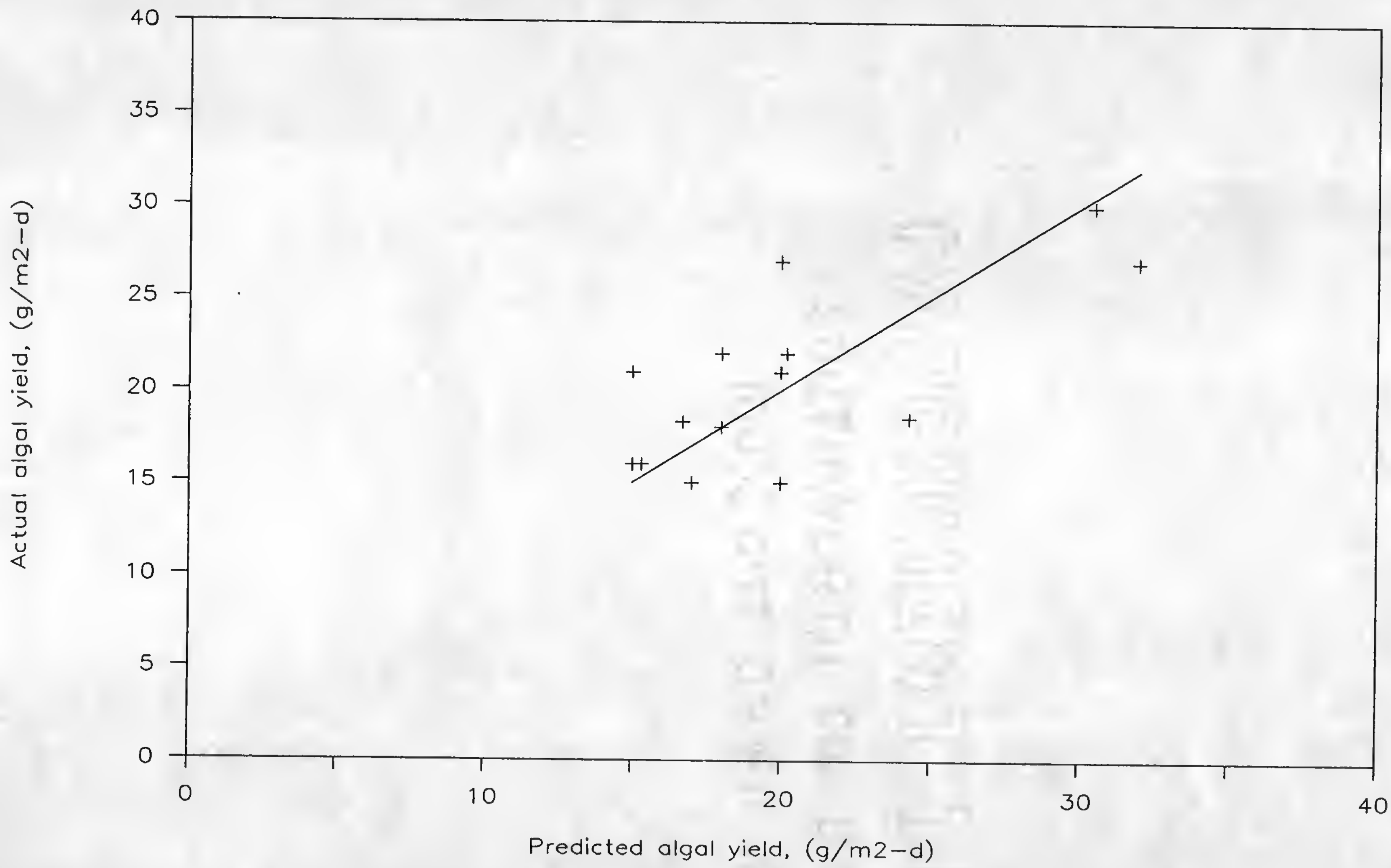

Figure 14. Predicted vs. actual algal yield.

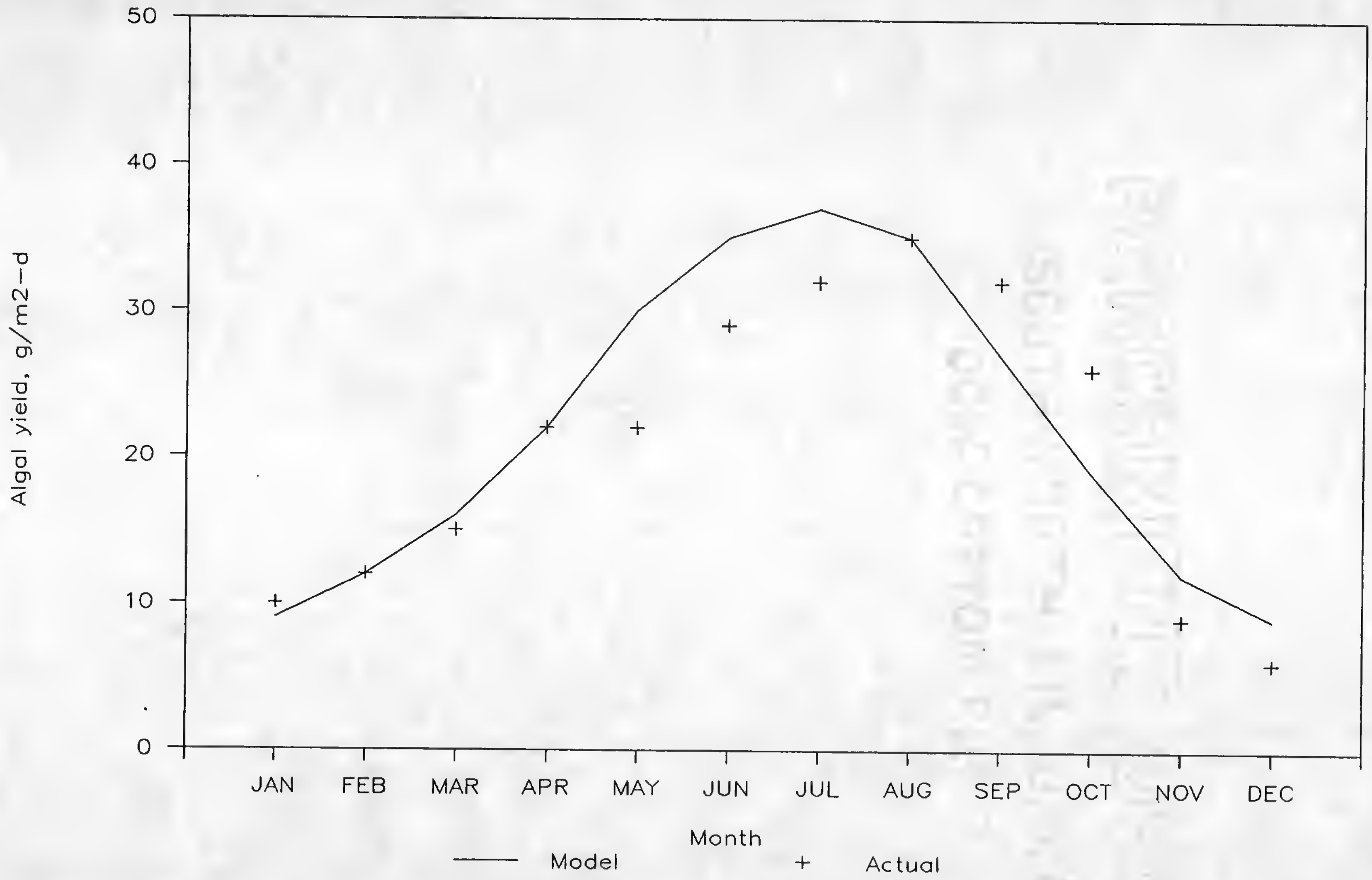

Figure 15. Simulation of algal growth in Haifa, Israel (Moraine et al., 1979).

curves may have been due to differences between actual conditions and the average climatological data used for the simulation.

Table 4. Algae model validation sources

Location	Source
Belfast, N. Ireland	Fallowfield and Garrett (1985)
Singapore	Goh and Lee (1982)
Haifa, Israel	Moraine et al. (1979)
Jerusalem, Israel	Shelef et al. (1973)
Haifa, Israel	Shelef et al. (1978)
Gainesville, Florida	Lincoln et al, (1986)
Roswell, New Mexico	Johnson et al. (1988)
Taiwan	Goldman (1979a)
Taiwan	Tsukada et al. (1977)
Tokyo, Japan	Tsukada et al. (1977)
Tokyo, Japan	Goldman (1979a)
Trebon, Czeckoslovakia	Goldman (1979a)
Rupite, Romania	Goldman (1979a)

Water Hyacinth Model

The water hyacinth model used the parameters originally specified by Lorber et al. (1984) with the exception of modification of the temperature function. As mentioned

earlier, the nitrogen and phosphate parameters were not used in this model.

The temperatures used for validation of the model were the actual monthly average air temperatures for Sanford, Florida for the year 1982. These temperatures were obtained from the 1982 United States Weather Bureau Climatological Data, Florida section. The temperatures used in the simulation are shown in Figure 16.

Solar insolation was simulated by a modified sine function. A comparison of the simulated data to the actual data from Orlando Florida is shown in Figure 17. The Orlando data is taken from tables presented by Lunde (1980). Considerable thunderstorm activity occurs in Florida in June and July. As a result, the insolation in central Florida during these months is somewhat less than would be expected at a similar latitude in other areas.

The model was calibrated with the data of Reddy (1983). A comparison of the model output with Reddy's data as presented by Lorber et al. (1984) is shown in Figure 18. It should be noted that the model underpredicted production at the end of the year. This may be attributed to the simulated solar radiation being lower than actual radiation at that time of year. Similarly, the model slightly overpredicted growth during the summer months when simulated radiation was higher than the actual radiation.

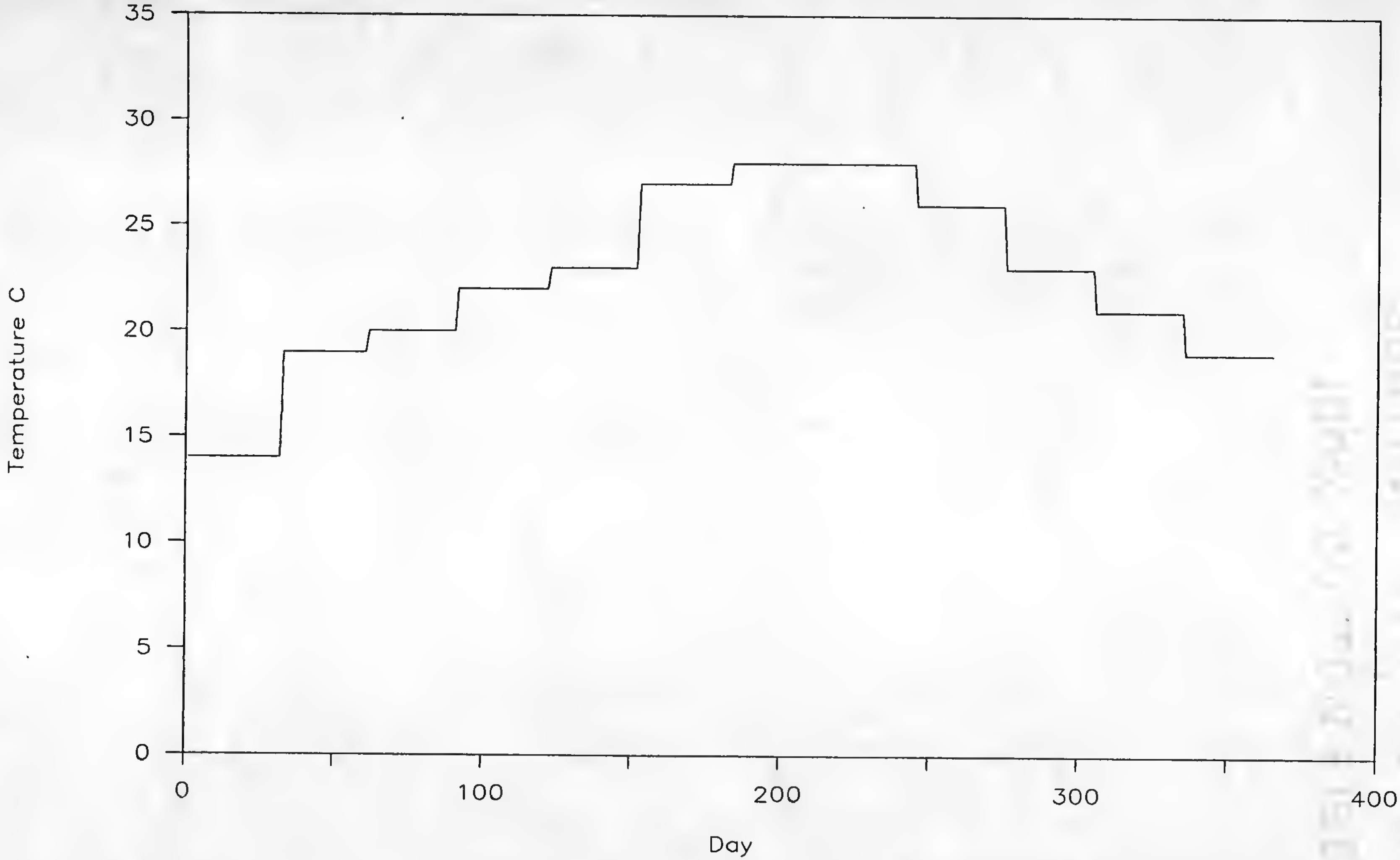

Figure 16. Temperatures used in simulation of water hyacinth growth data of Reddy (1983) from Sanford, Florida.

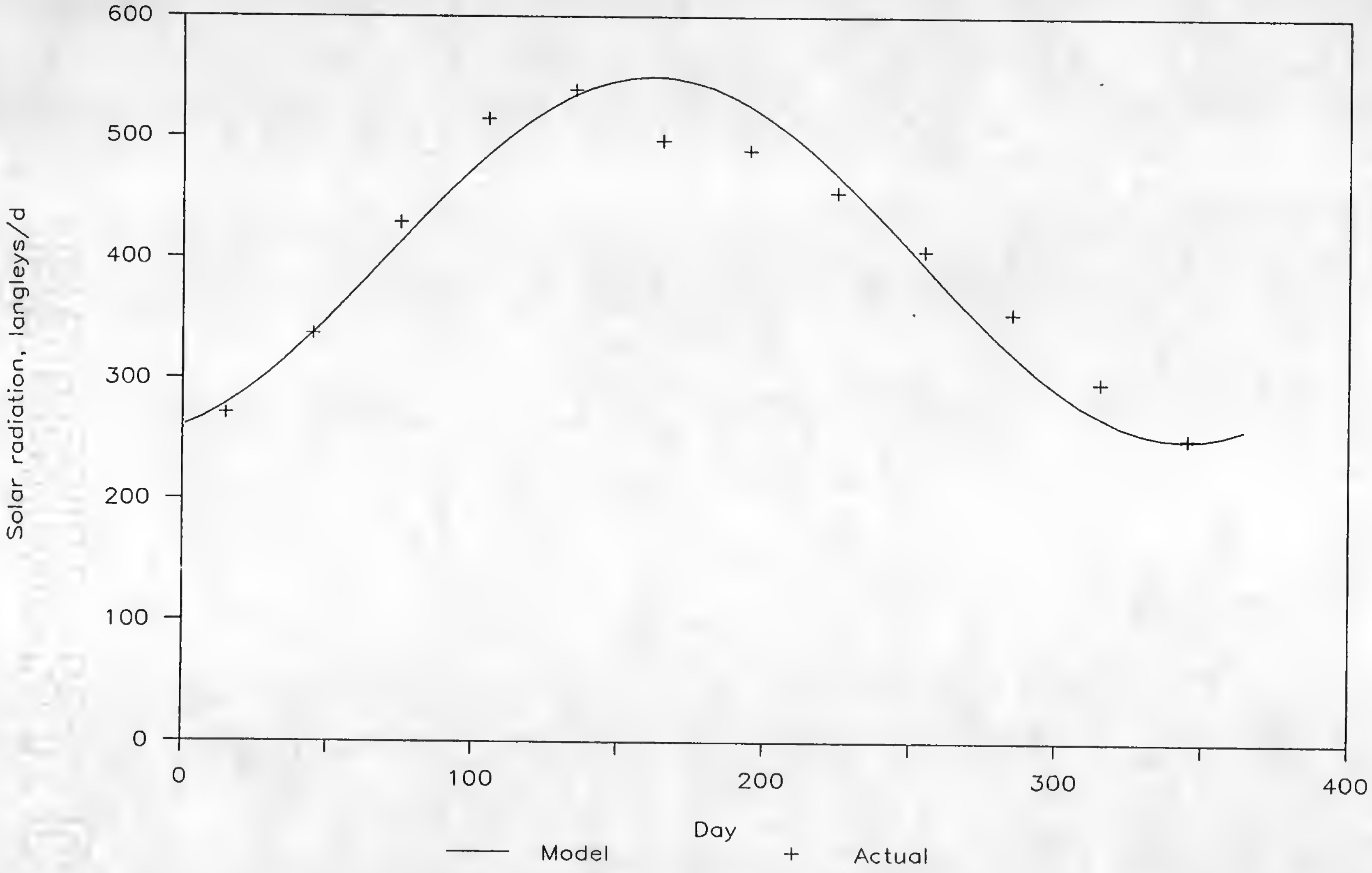

Figure 17. Simulated and actual daily solar insolation in Orlando, Florida.

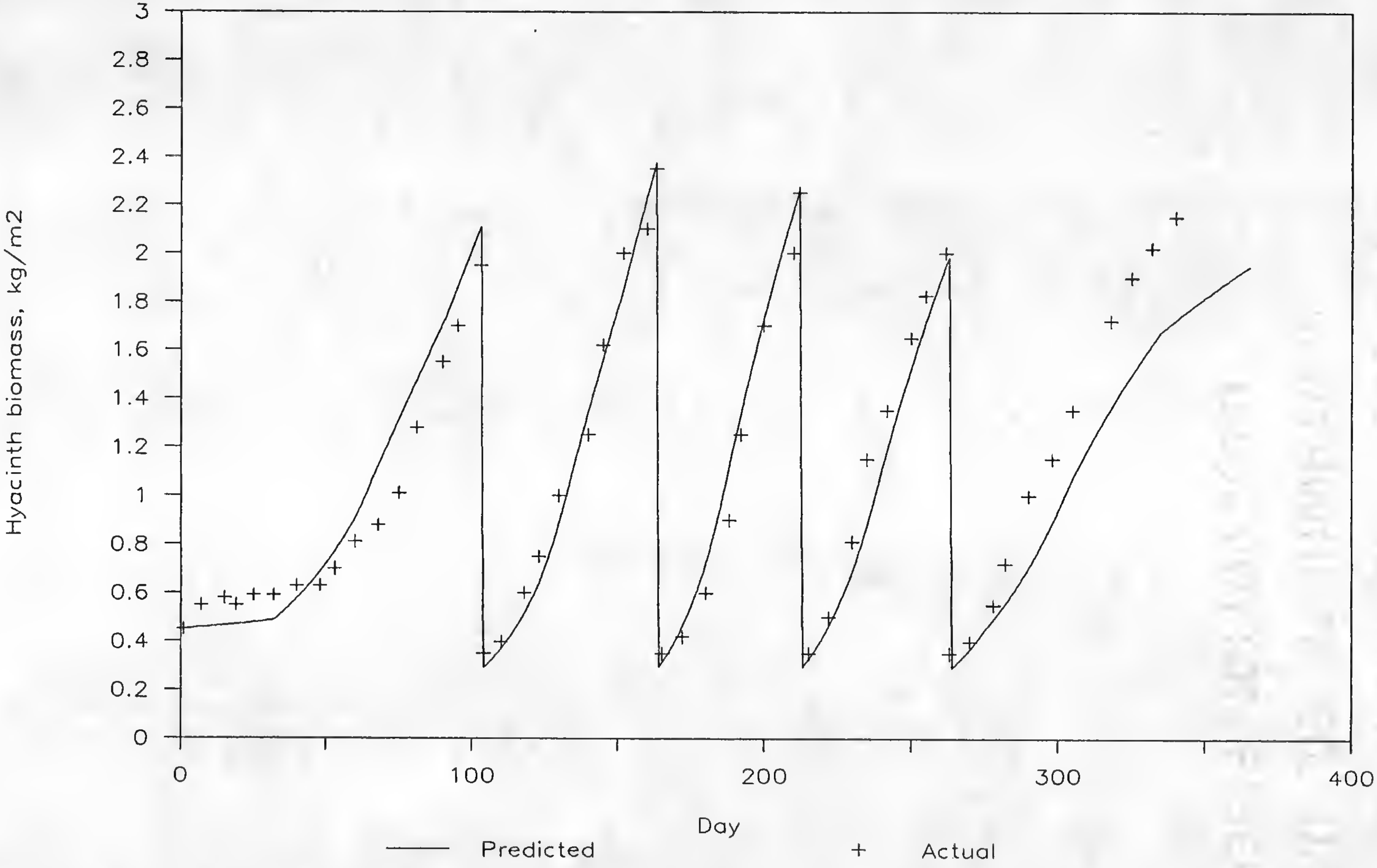

Figure 18. Simulation of data from Reddy (1983).

The model was validated by comparison with growth rate data from Debusk et al. (1981), Reddy et al. (1985), Reddy and Debusk (1984), and Debusk and Reddy (1987). The model closely approximated the data from 1981, 1984 and 1987 (Figure 19), when the hyacinths were grown in nutrient rich environments. Sharp drops in model productivity are due to the effects of harvesting. Data from Reddy et al. (1985) were from hyacinths grown in channels fed by secondary sewage effluent, and they may have been nutrient limited. Actual data from the above sources was plotted against model predictions and is shown in Figure 20. The statistical test for lack of fit was not significant.

The average growth rate of 24.9 g/m^2-d when simulating growth in central Florida compared with 23.2 g/m^2-d reported by Joglekar and Sonar (1987) for their work in India. The maximum growth rates were within the 50 to 64 g/m^2-d range reported by other investigators (Debusk et al., 1981; Reddy and Debusk, 1984).

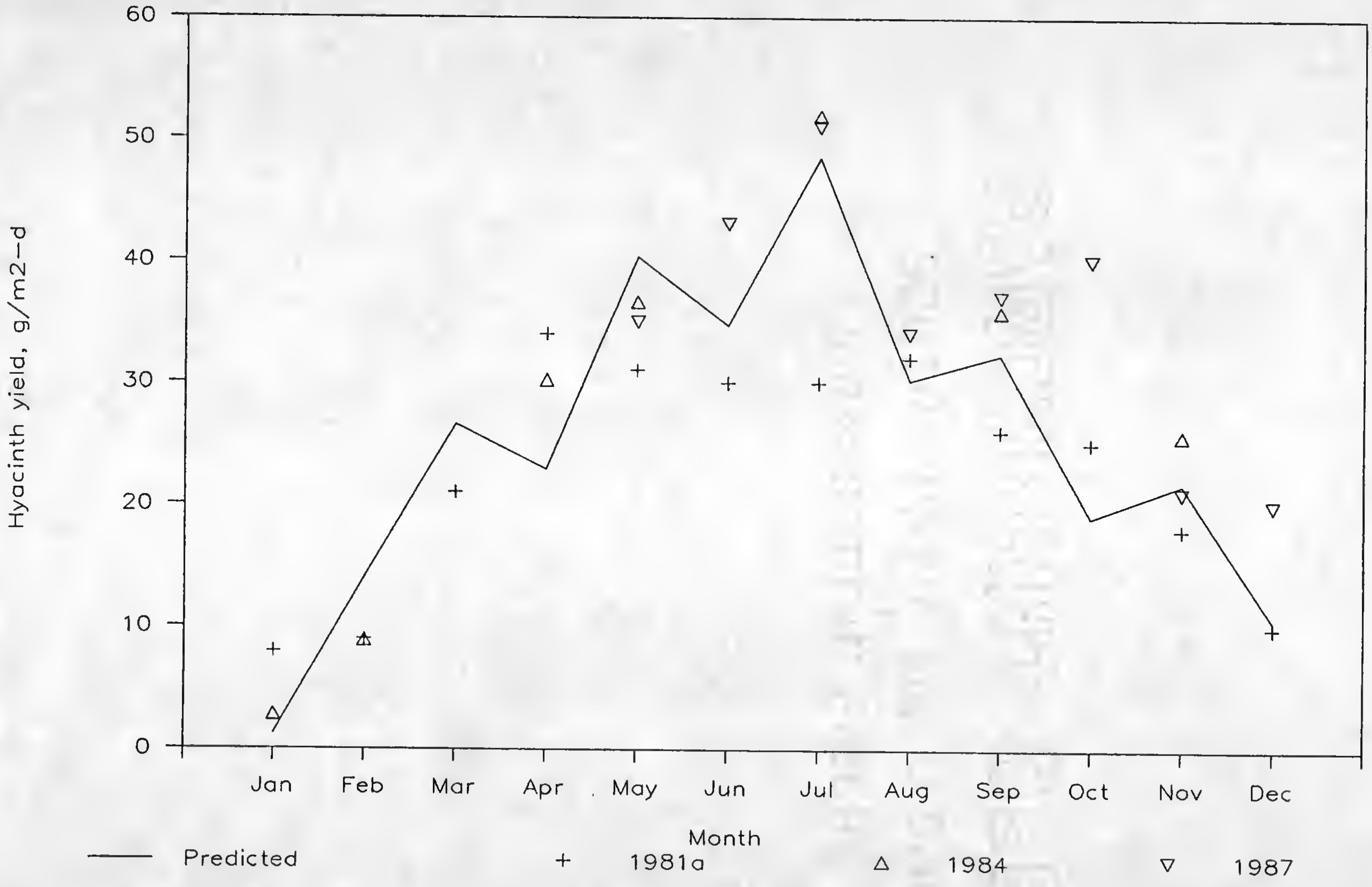

Figure 19. Model predictions and average monthly yields for three studies of water hyacinth growth in central Florida.

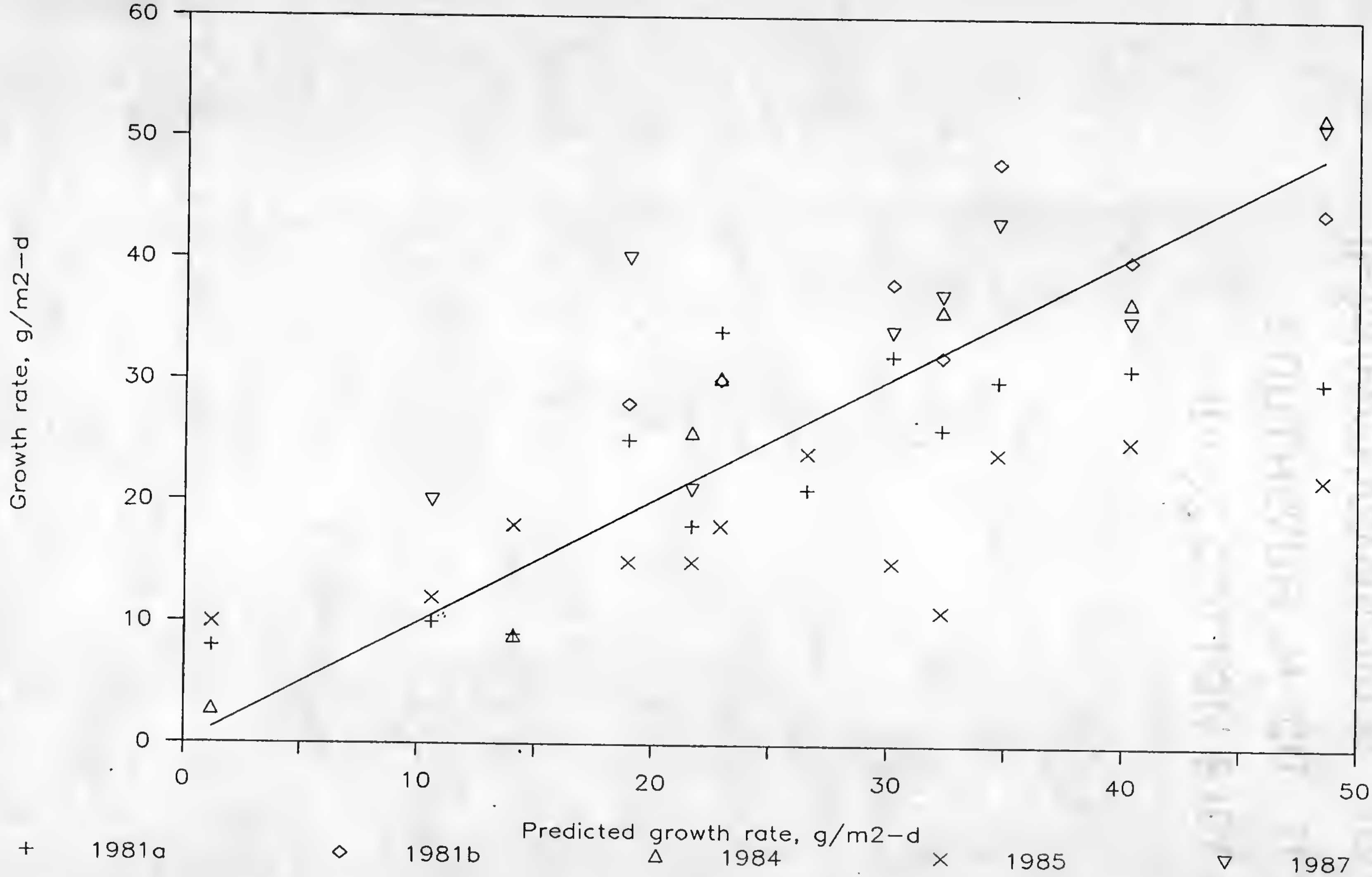

Figure 20. Predicted vs. actual water hyacinth growth rates for five studies in Florida.

RESULTS AND DISCUSSION

The model was coded in Turbo Pascal 3.0 (Borland International Inc. Scotts Valley, CA) for use on personal computers. The initial operating conditions used for system input are listed in Table 5. A more detailed explanation of the system variables and their nomenclature is located in Appendix B. The sunlight and temperature functions were based upon sine functions. A random number generator was used to modify the otherwise smooth curve to account for daily fluctuations. A flow chart outlining basic program operation is provided in Appendix C.

The anaerobic digestion component of the model was found to provide an excellent vehicle for the simulation of disparate reactor types by modifying only the calling procedure. In addition, this section of the model has demonstrated the ability to predict digester operation with a variety of feedstocks using only a few parameters to characterize those feedstocks. This has profound implications for the researcher who until now has attempted to create individual models for each type of digester and for each substrate. This practice has resulted in vast expenditures of time and energy which have frequently gone for

Table 5. Base input values for simulation variables

Variable	Value	Units
algarea	2	ha
algchmreq	0.01	kg / m^3
algenreq	0.079	MJ / m^3
algfxlbreq	4	manhours / harvest
alghrvfrq	1	day
alghrvlim	400	mg/L
alglolim	250	mg/L
algvalue	400	$ / tonne
algeffdepth	0.3	m
algvrlbreq	0.5	manhours / tonne
blrvalue	0.00476	$ / MJ
boilreff	86	%
boilrsize	0	MJ
chemcst	0.15	$ / kg
chopsize	6.35	mm
depryers	30	years
dmpflcst	0	$ / m^3
dmpvscst	0	$ / tonne
endday	365	days
enrindx	4457	
gastrcap	250	m^3
geneff	20	%
genhtval	0.00476	$ / MJ
gensize	4320	MJ / day
hyalolim	1000	mg/L
hyarea	1	ha
hyenreq	20	MJ / tonne
hyhrvfrq	7	days
hyhrvlim	2400	mg/L
hylabreq	0.8	manhours / tonne
hysolids	12	%
hyvsofts	50	%
intrstrt	9	%
laborcst	7	$ / hour
landcost	150000	$
powercst	0.021	$ / MJ
pwrvalue	0.014	$ / MJ
rx1brc	0.995	
rx1hx	75	%
rx1ktot	5	
rx1mint	32	degrees C
rx1rval	2.817	m^2-sec.-deg.C /J
rx1vol	80	m^3
rx2brc	1	
rx2ktot	1	

Table 5 (continued)

Variable	Value	Units
rx2maxvol	500	m^3
rx2sedvs	30	mg/L
rx2seed	0.2	
rx3brc	0	
rx3hx	65	%
rx3ktot	1	
rx3mint	35	degrees C
rx3mxhrs	4	hours
rx3rval	1.937	m^2-sec.-deg.C /J
rx3vol	150	m^3
wasthteff	40	%
flow	30	m^3/d
VS input	30	mg/L
FBR % flow ($f\%_1$)	70	%
CER % flow ($f\%_2$)	2	%
CSTR % flow ($f\%_3$)	28	%
FBR VS % ($VS\%_1$)	42	%
CER VS % ($VS\%_2$)	2	%
CSTR VS % ($VS\%_3$)	56	%
average sunlight	16.72	MJ / m^2-day
lowest avg. monthly temp	14	degrees C

naught when the feedstock or the digester configuration was changed.

Estimates of utility and labor costs, energy usage and unit efficiency were taken from a variety of sources and should not be assumed to represent any particular system. Variations of an order of magnitude may not be unreasonable in some cases, depending on the physical nature of harvest operations, mixing methods, etc.

System simulations were performed to test the effects of various algal harvest frequencies, sale prices of algae, gas usage practices, flow management practices, climates, system sizes and substrate types on the system economics.

The basic system used for simulation was assumed to have an algal production area of 2 ha. The algal product was assumed to be harvested at the beginning of the day after reaching an algal density of 400 g/m^3. Optionally, algae could be harvested on a regular basis if high densities were not reached. The impact of various harvesting strategies on the cost of algal production is shown in Figure 21. As may be seen, the minimum amount of labor required to perform a harvest, regardless of the amount harvested, leads to the conclusion that harvest schedules for the algae should be dependant upon the algal density. The most economical method was to allow for harvesting only when a maximum density was reached. This reduced unnecessary labor expense during slow growth periods.

The effect of the sales price of harvested algae on net operating revenues is shown in Figure 22. Because algal production is a such a large part of the system as modeled, the use for which algae may be sold is of considerable importance. If synthetic oil from lipids is the primary use, the revenues may be higher than if the algae were used for animal feed.

The was used for generating electrical power or for generating steam from a boiler. When electrical power was generated by a conventional internal combustion engine the cooling water was used to provide hot water in lieu of purchasing energy for hot water heating. If no provisions

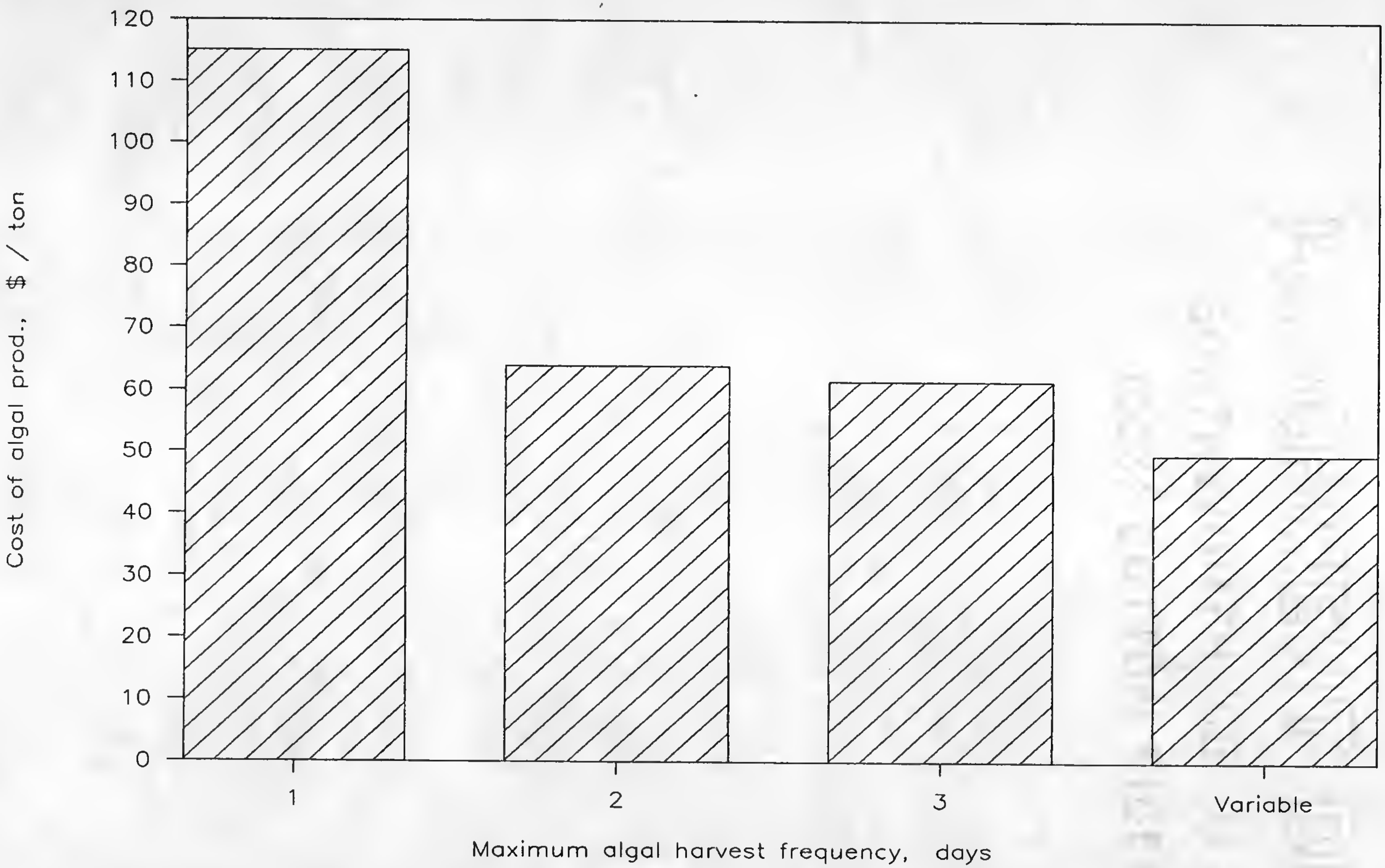

Figure 21. Unit cost of algae as a function of the maximum time between harvests. "Variable" indicates harvesting only when the density limit is exceeded.

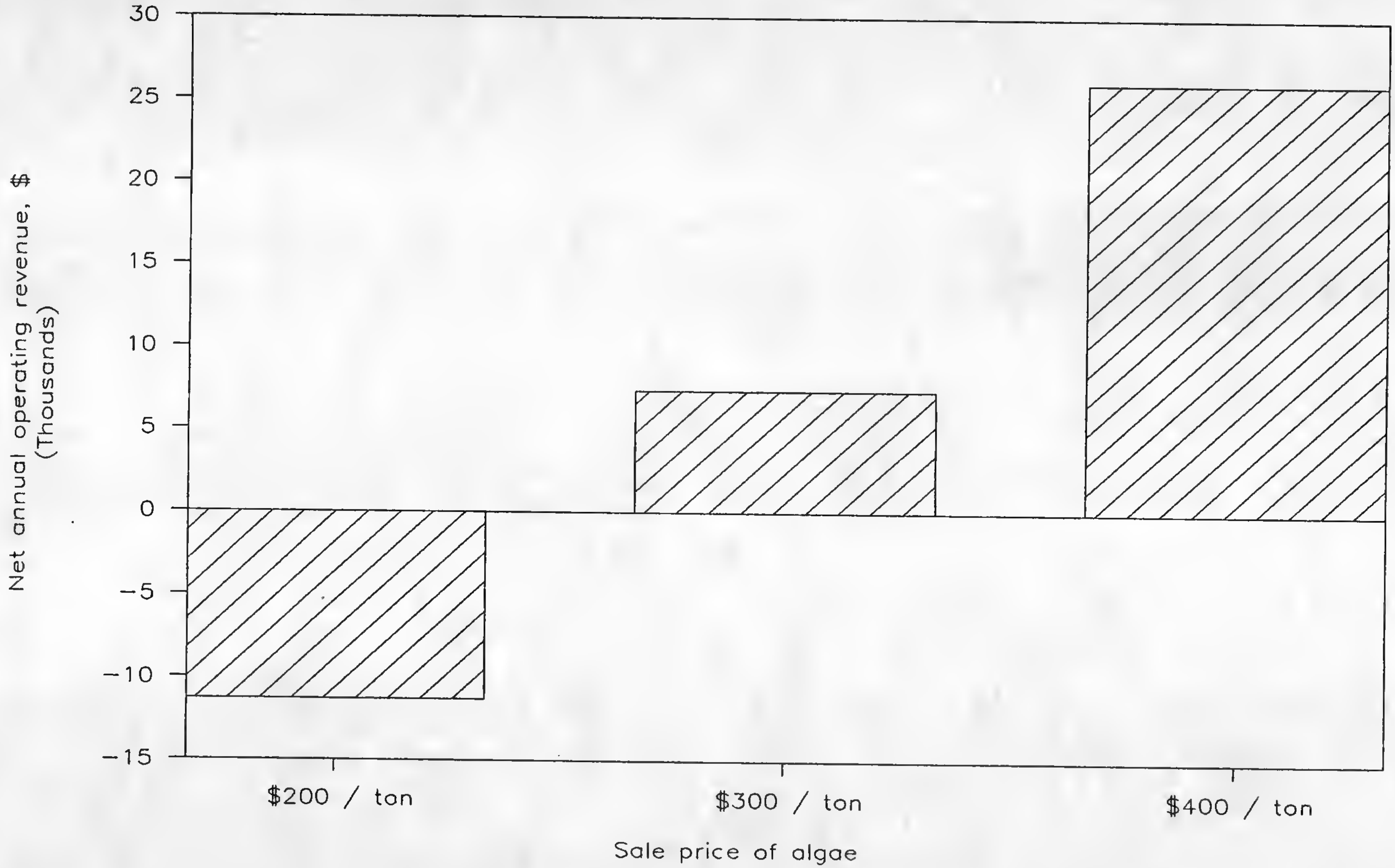

Figure 22. Net operating revenue for three values of algae. Price is in dollars per ton dry weight.

were made for the hot water to be used the value of the retained heat was zero. The effects of various biogas management strategies on net operating revenue is shown in Figure 23. The results shown depend heavily upon the exact values chosen for electrical power cost vs. wholesale value, steam value, and the value of the heated water from the generator cooling system.

As with any system designed to produce energy products, the overall economic outlook is heavily dependant upon world energy prices.

The type of substrate used for digestion may also have an effect on the bottom line. Simulations were carried out using swine, beef, poultry, and dairy wastes as the substrate. The effects on net operating revenue are shown in Figure 24. As would be expected, the results are roughly proportional to the biodegradability of the material.

The climate is a significant factor in almost any agricultural operation. Systems simulations were performed, using annual average temperatures and amounts of sunlight, for three locations in the United States. These locations were Gainesville, Florida; Knoxville, Tennessee; and Yuma, Arizona. As might be expected, the colder climate in Tennessee was less favorable to system operation as is shown in Figure 25. Algal production and water hyacinth growth are most suited to tropical and subtropical climates.

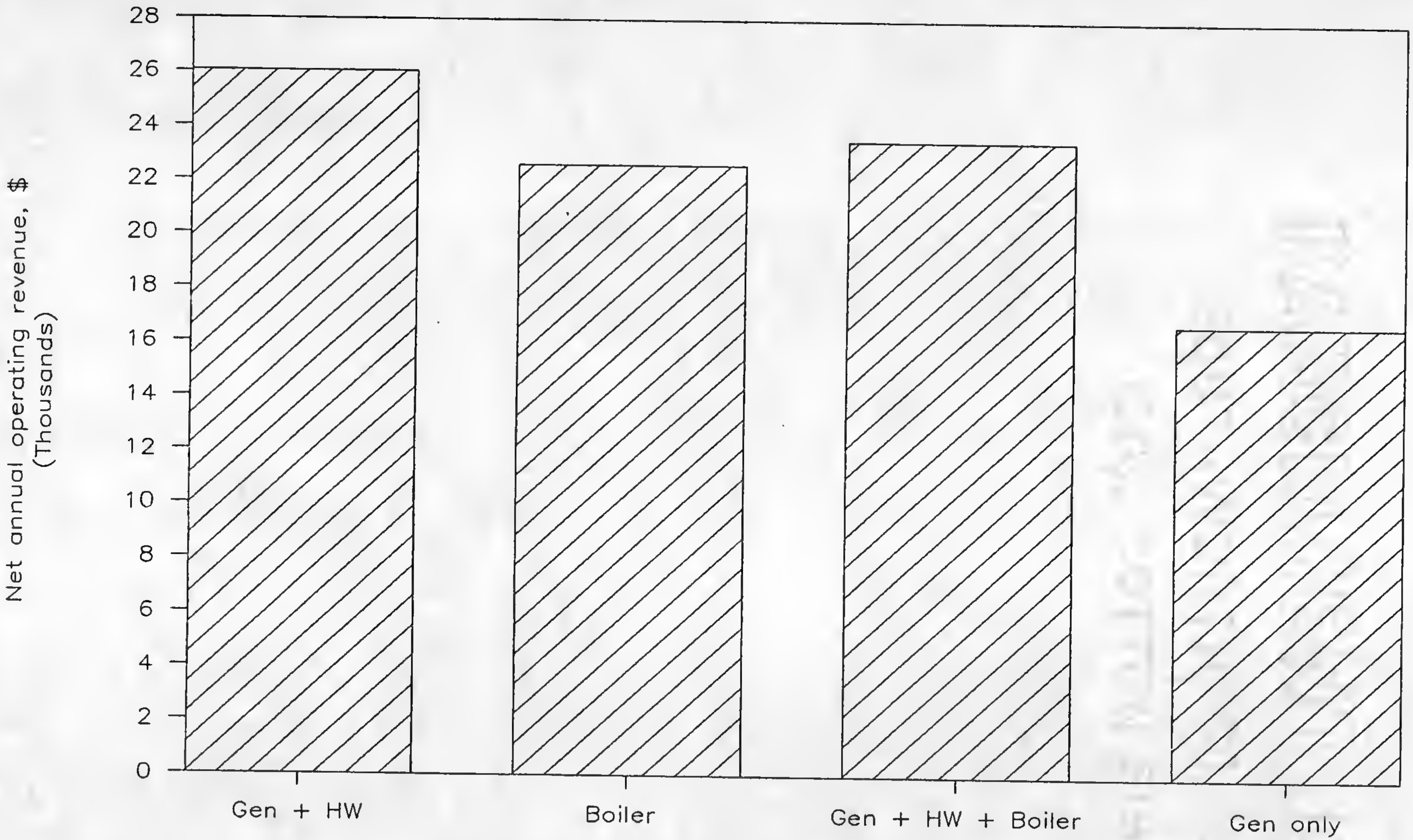

Figure 23. Net operating revenue as function of biogas usage. Hot water (HW) may be obtained from the generator cooling system.

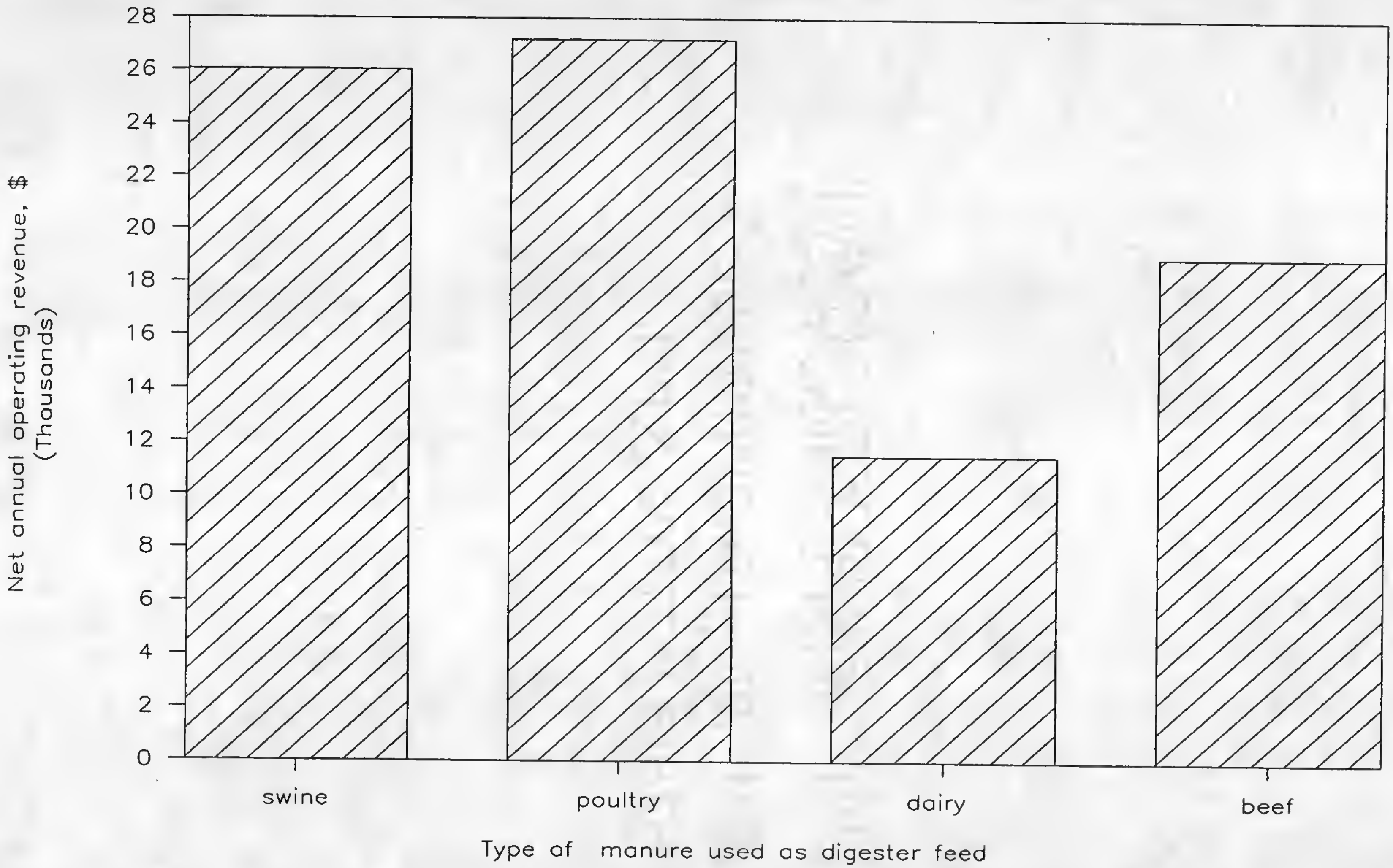

Figure 24. Net operating revenue as a function of digester substrate.

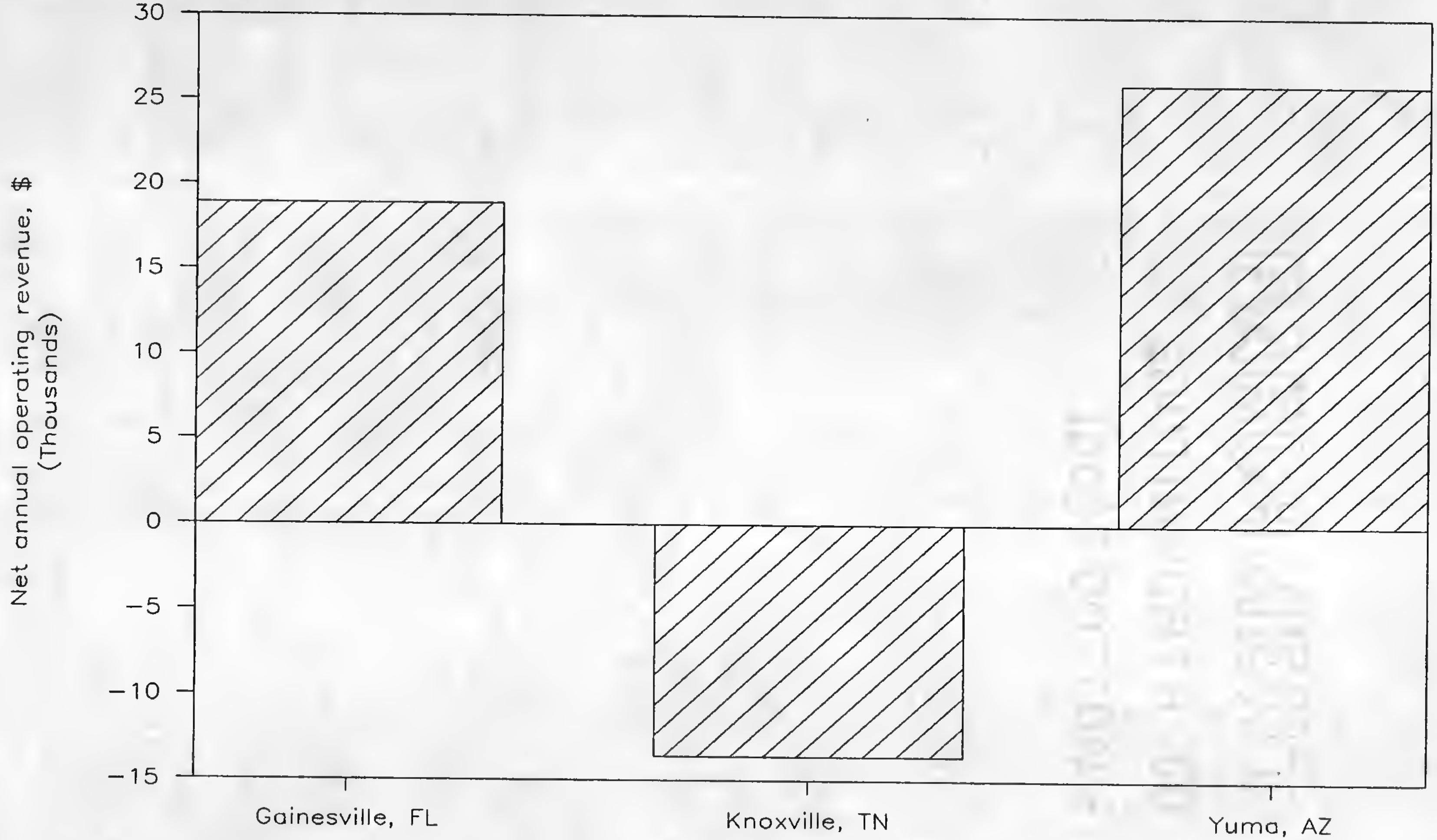

Figure 25. Net operating revenue as a function of climate. Simulation used average annual temperatures and solar insolation.

Simulation of net operating revenues for one year for a hypothetical system as in Table 5 and located in central Florida are shown in Figure 26. The average daily revenue from the system was $44.02.

Flow management was another area examined in light of its effects on the bottom line. Net annual revenues for the system are shown in Figure 27 for three flow schemes. In these simulations the 70%-28% FBR-CSTR flow distribution represents the basic flow scheme used in Table 5. The CER was allocated 2% of the influent hydraulic flow and VS in all simulations. For the simulation of a 49-49 or 24-74 FBR-CSTR distribution the VS distribution was the same as the flow distribution. The shorter retention time permitted by the FBR, which result in a smaller vessel and generally a lower capital cost, is reflected by the greater (less negative) net revenue in Figure 27.

Offseting this is the inability of the FBR to handle high solids waste without clogging. The choice of reactor types to be used in a system is dependant on a variety of factors in addition to capital cost.

Efficiencies of scale were evaluated by expanding the basic system described in Table 5 by factors of ten and one hundred. The fixed labor requirement for algal harvest was also increased. The results are shown in Figures 28-34. As may be seen from Figure 28, the probable break even point was estimated to be in excess of 3785 m^3/d. The land area

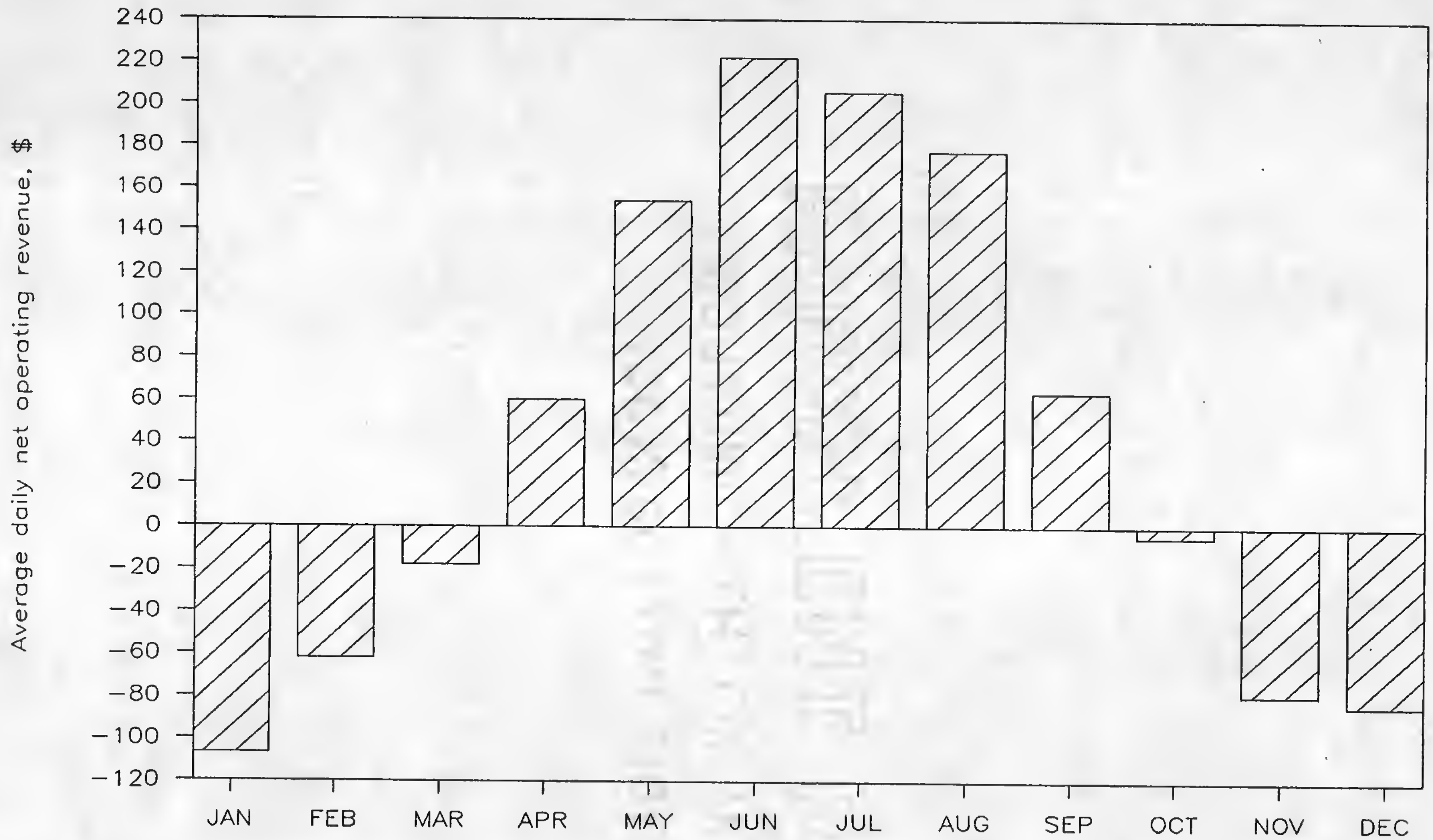

Figure 26. Average daily net operating revenue as a function of time for a typical system in central Florida.

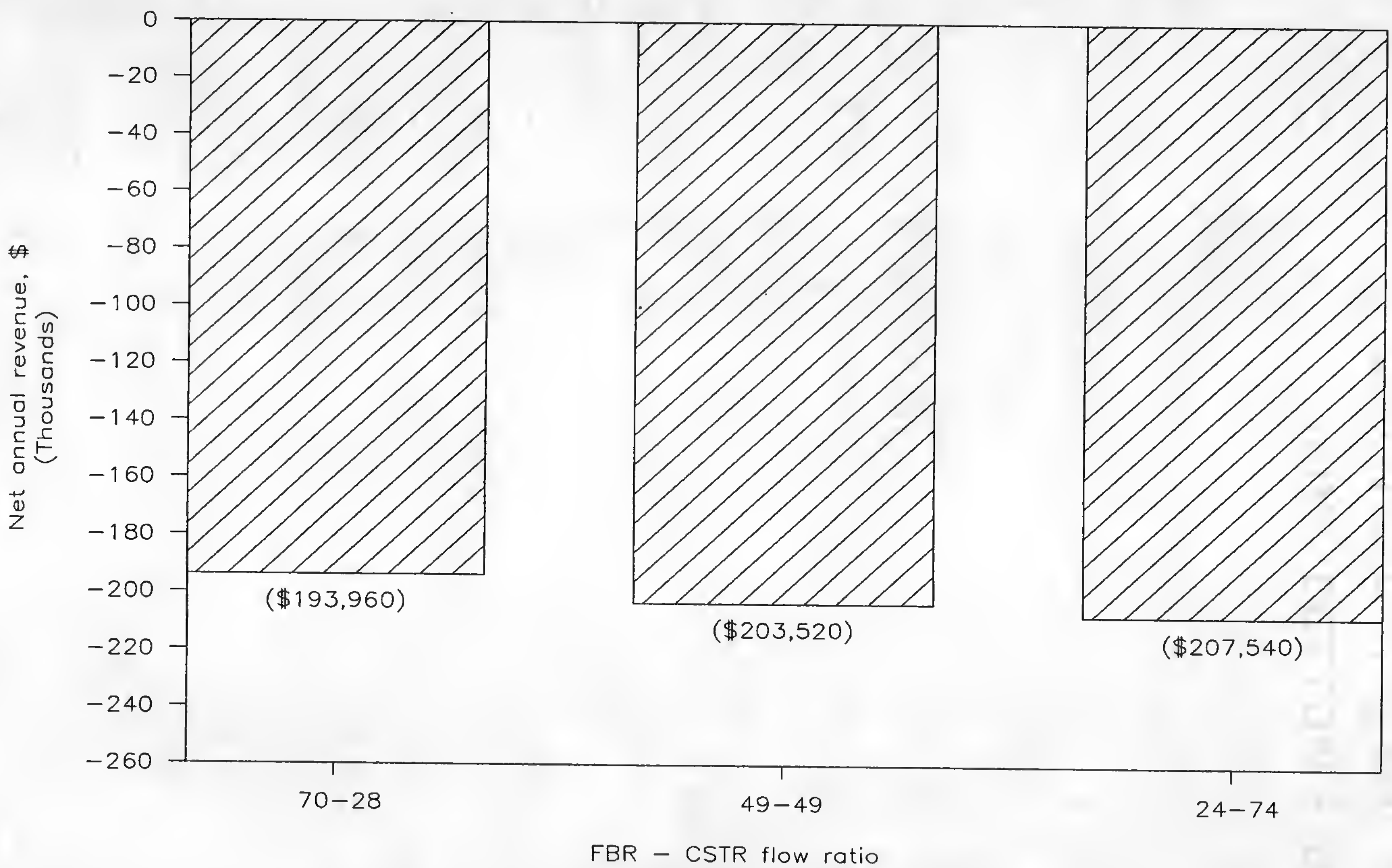

Figure 27. Net annual revenue for three flow management schemes. The CER receives 2% of the flow.

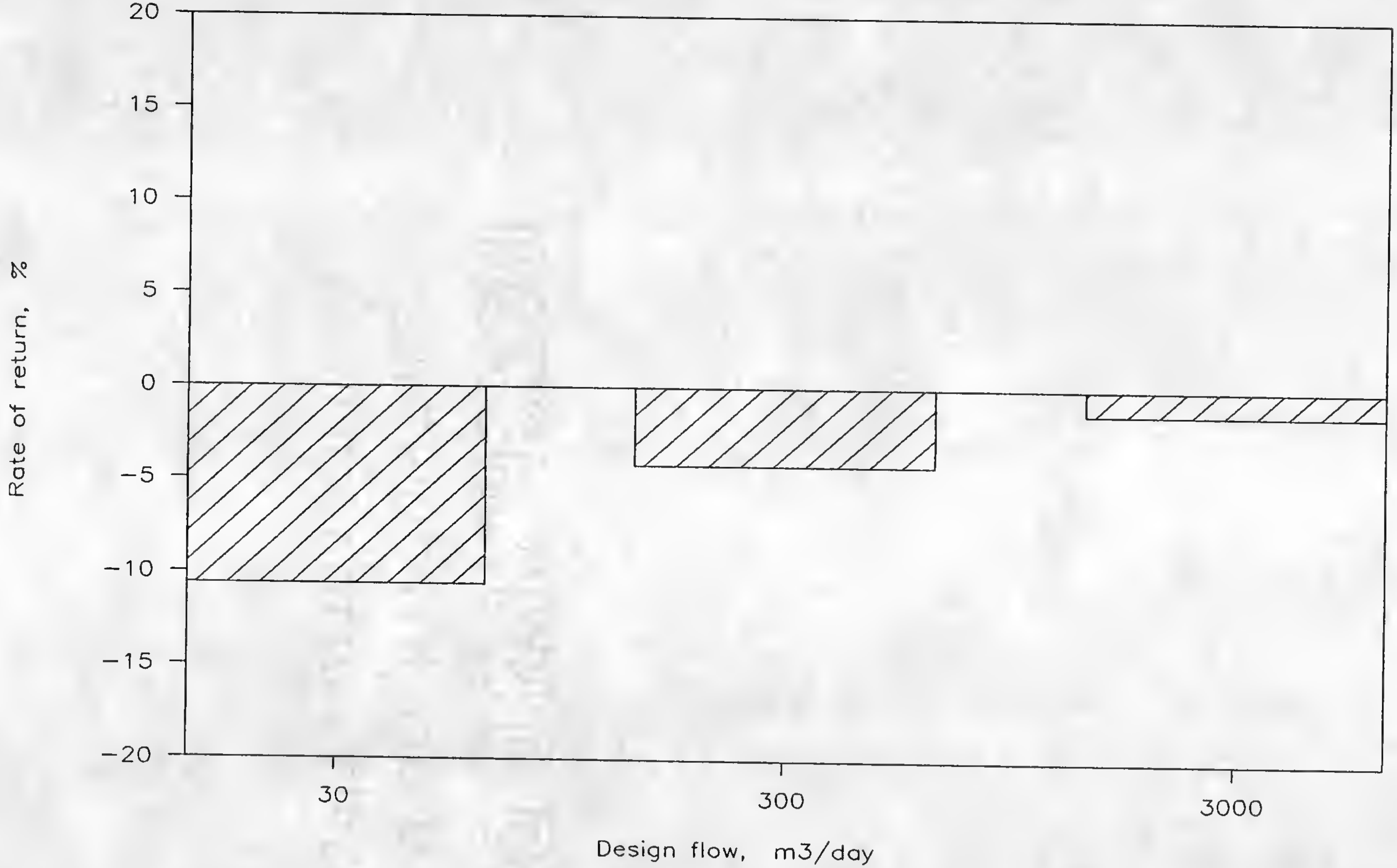

Figure 28. Rate of return as a function of system size.

devoted to algal production would approach 259 ha. Algae is the primary source of revenue in the system described, and the distribution of revenue is relatively constant with variations in system size (Figures 29 and 30).

The distribution of the operating expenses was determined to vary with system size. For the very small system (30 m^3/d) the dominant cost was that of general system labor (Figure 31). In the larger system (3000 m^3/d) the fixed labor requirement was overshadowed by the variable labor requirements associated with product harvest (Figure 32). This was to be expected as a minimum of three full time employees was assumed regardless of system size.

In addition to the change in the distribution of operating costs, the relative percentage of operating costs as a part of the total annual cost also decreases as the size of the system increases. In the smaller system (30 m^3/d) operating costs make up 24.6% of the total annual cost of the system (Figure 33). This percentage drops to 7.5% for the 3000 m^3/d system (Figure 34). This reinforces the notion of economies of scale. Figures 33 and 34 also emphasize the effect that interest rates will have on proposed construction of such a system.

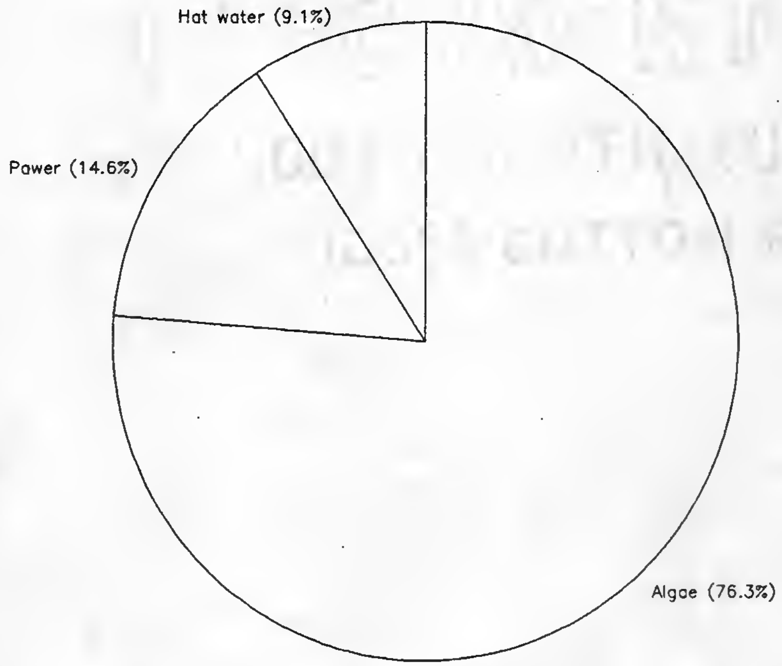

Figure 29. Revenue distribution for a 30 m^3/d system.

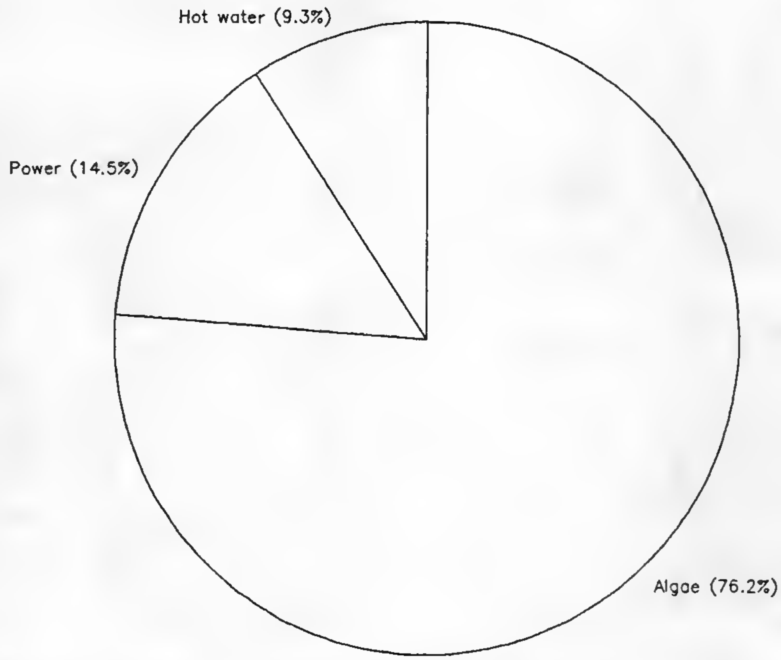

Figure 30. Revenue distribution for a 3000 m^3/d system.

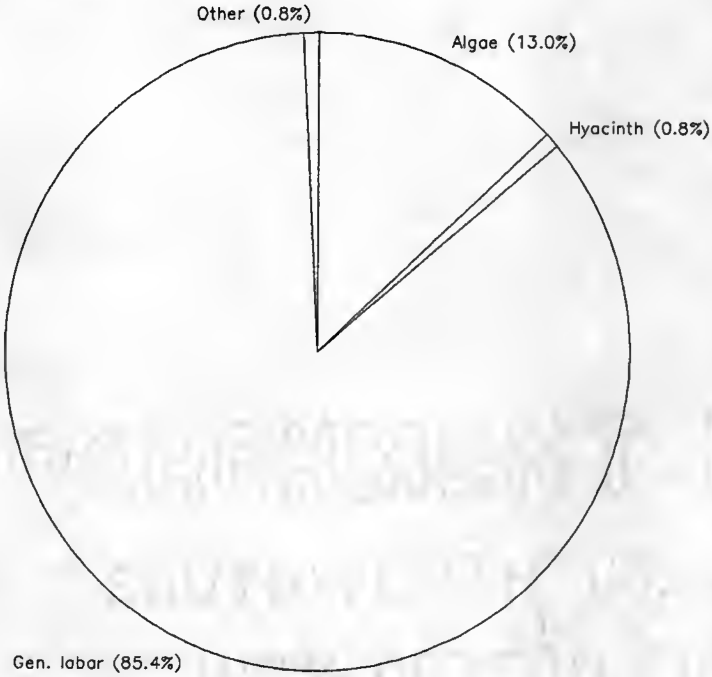

Figure 31. Operating expense allocation for a 30 m^3/d system.

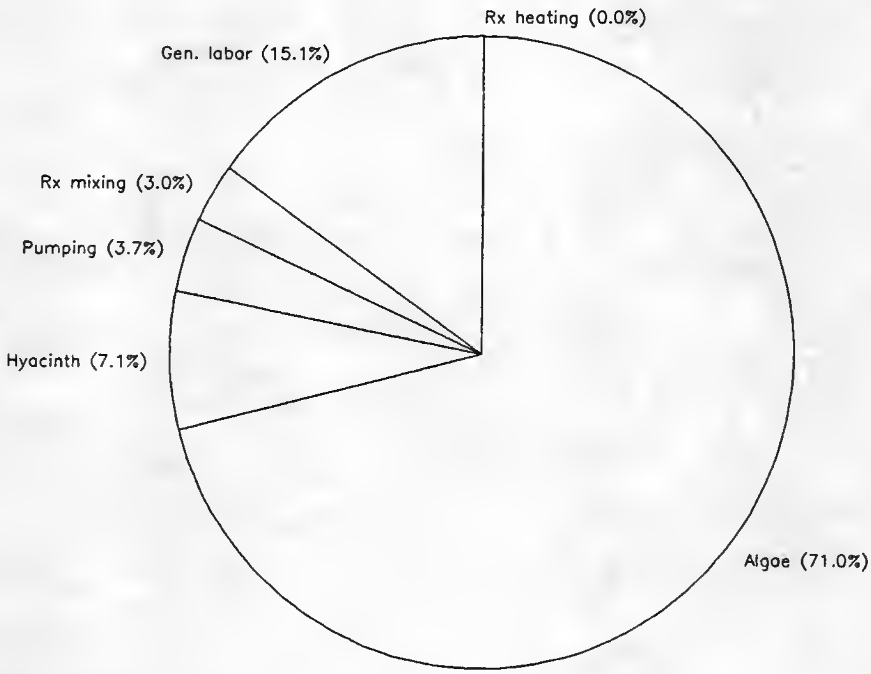

Figure 32. Operating expense allocation for a 3000 m^3/d system.

Figure 33. Cost distribution for a 30 m^3/d system.

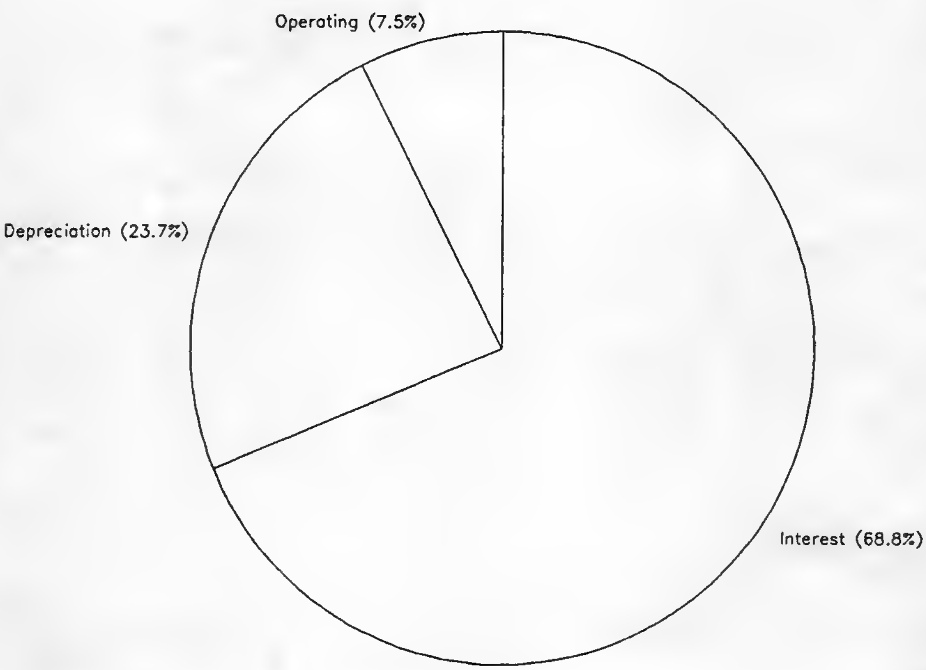

Figure 34. Cost distribution for a 3000 m^3/d system.

CONCLUSIONS

1. A mathematical model of an integrated system was developed with the ability to assist in the determination of design parameters and management practices by providing relative economic information for various design and operating strategies.

2. A generic anaerobic digestion submodel was developed which provided a tool for the testing of conversion efficiencies and kinetic parameters for a wide range of substrate types and reactor designs.

3. Simulations using the system model have shown that when digested anaerobically poultry and swine manure provide greater biogas yields than beef or dairy manure.

4. An integrated system using algae and water hyacinths was found to be best suited to tropical or semi-tropical locations with warm temperatures and high annual solar insolation.

5. The availability of large quantities of digestible waste and low land prices were found to be desirable in order to take advantage of the economies of scale.

6. It was determined that algal harvesting should be performed using a density dependant strategy as opposed to a fixed schedule in order to minimize labor costs.

RECOMMENDATIONS FOR FURTHER RESEARCH

1. The system model should be expanded to track nitrogen, phosphorus, and pH throughout the system.

2. Additional experimental work is needed to quantify the effects of anaerobic digestion on the availability of inorganic nutrients to downstream processes.

3. The anaerobic digestion model should be expanded to include the four currently recognized bacterial groups. This will also enable pH estimation and facilitate estimates of inorganic nutrient availability.

4. An algal pond model should be included to account for nutrient losses through volatilization. The algal growth model should be expanded to include nitrogen and phosphorus.

5. The hyacinth model should be expanded to respond to N and P levels. Toxic effects, many of which may be unknown, and freezing should be added.

6. Additional work is required to understand the nutrient cycle in hyacinth ponds. Uptake by hyacinths alone is insufficient to accurately model pollutant removal through a pond system.

7. Additional work is required to characterize the breakdown process of complex organics in the digester

environment. The effects of bacterial concentrations, growth rates, and enzyme levels should be determined.

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APPENDIX A

SYSTEM FLOW VARIABLES

f_1	Flow into the FBR.
f_2	Combined flow into the CER.
f_{2l}	Flow into the CER from the influent substrate.
f_{2h}	Flow into the CER from the hyacinth harvest.
f_3	Flow into the CSTR.
f_a	Flow into the algal pond.
f_{aout}	Flow out of the algal pond.
f_{dl}	Influent flow diverted to the system dump.
f_{d2}	Digester effluent flow diverted to the system dump.
f_{d3}	Algal effluent flow diverted to the system dump.
f_h	Flow into the hyacinth ponds.
f_{ha}	Algal effluent flow into the hyacinth ponds.
f_{hr}	Digester effluent flow into the hyacinth ponds.
f_{in}	System influent flow.
$f\%_1$	Fraction of system influent flow which goes to the FBR.
$f\%_2$	Fraction of system influent flow which goes to the CER.
$f\%_3$	Fraction of system influent flow which goes to the CSTR.

$f\%_{a1}$	Fraction of FBR effluent flow which goes to the algal unit.
$f\%_{a3}$	Fraction of CSTR effluent flow which goes to the algal unit.
$f\%_{h1}$	Fraction of FBR effluent flow which goes to the hyacinth unit.
$f\%_{h3}$	Fraction of CSTR effluent flow which goes to the hyacinth unit.
S_1	Volatile solids (VS) entering the FBR.
S_{21}	VS from the system influent entering the CER.
S_{2h}	VS from the harvested hyacinths entering the CER.
S_3	VS entering the CSTR.
S_a	VS entering the algal unit.
S_{aout}	VS leaving the algal unit in the effluent flow.
S_{d1}	Influent VS diverted to the system dump.
S_{d2}	Digester effluent VS diverted to the system dump.
S_{d3}	Algal effluent VS diverted to the system dump.
S_h	VS entering the hyacinth unit.
S_{ha}	VS entering the hyacinth unit from the algal unit.
S_{hr}	VS entering the hyacinth unit from the reactors.
s_1	Concentration of volatile solids (VS) entering the FBR.
s_{21}	Concentration of VS from the system influent entering the CER.
s_{2h}	Concentration of VS from the harvested hyacinths entering the CER.

s_3	Concentration of VS entering the CSTR.
s_a	Concentration of VS entering the algal unit.
s_{aout}	Concentration of VS leaving the algal unit in the effluent flow.
s_{d1}	Concentration of VS diverted to the system dump from the influent.
s_{d2}	Concentration of VS diverted to the system dump from the digester effluent.
s_{d3}	Concentration of VS diverted to the system dump from the algal effluent.
s_h	Concentration of VS entering the hyacinth unit.
s_{ha}	Concentration of VS entering the hyacinth unit from the algal unit.
s_{hr}	Concentration of VS entering the hyacinth unit from the reactors.
s_{in}	Concentration of VS in the system influent.
$s\%_1$	Fraction of system influent VS which goes to the FBR.
$s\%_2$	Fraction of system influent VS which goes to the CER.
$s\%_3$	Fraction of system influent VS which goes to the CSTR.
$s\%_{a1}$	Fraction of FBR effluent VS which goes to the algal unit.
$s\%_{a3}$	Fraction of CSTR effluent VS which goes to the algal unit.

$s\%_{h1}$	Fraction of FBR effluent VS which goes to the hyacinth unit.
$s\%_{h3}$	Fraction of CSTR effluent VS which goes to the hyacinth unit.

APPENDIX B

SYSTEM INPUT VARIABLES

algarea	The area of the algal growth ponds.
algchmreq	The amount of chemical flocculating agent required for algal harvesting.
algenreq	The amount of mixing energy required for algal production.
algfxlbreq	The amount of labor required per algal harvest, regardless of the amount harvested.
alghrvfrq	The frequency at which algae will be harvested regardless of the density.
alghrvlim	The density at which algae will be harvested, regardless of the day.
alglolim	The density of algae after harvest.
algvalue	The sales price of the dried algae.
algeffdepth	The effective depth of the algal pond.
algvrlbreq	The amount of labor required per tonne of algae harvested.
blrvalue	The value of steam from the boiler.
boilreff	Boiler efficiency.
boilrsize	The number of hp-hours produced by the boiler in 24 hours if operated at rated capacity.
chemcst	The cost of flocculating chemicals.

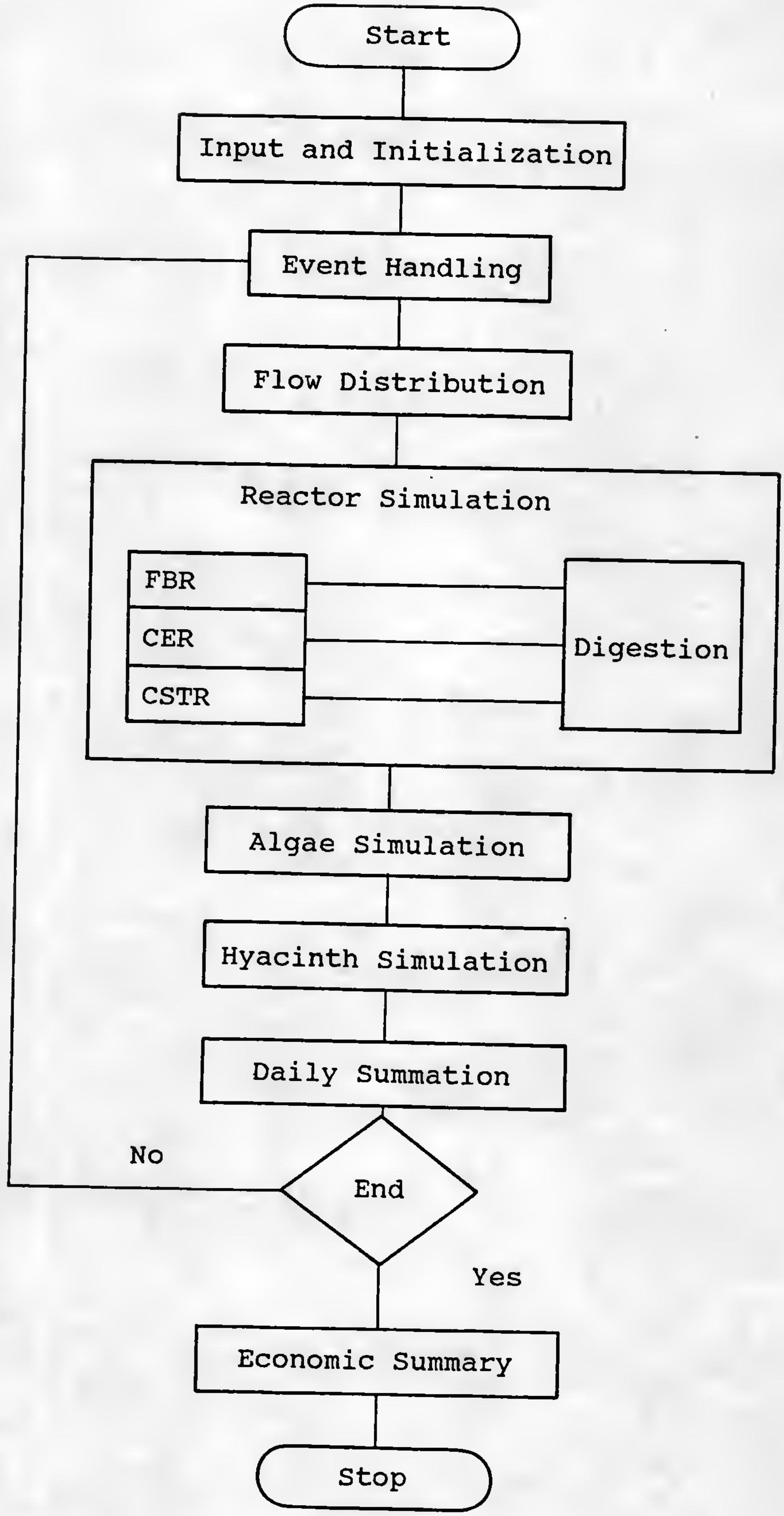
chopsize	The size to which hyacinths are reduced prior to feeding to the CER.
depryers	Depreciation period.
dmpflcst	Volumetric cost of waste treatment.
dmpvscst	Waste treatment surcharge for volatile solids (may be correlated to BOD, COD).
endday	Number of days to be simulated.
enrindx	Engineering News Record Construction Cost Index.
gastrcap	Biogas storage capacity.
geneff	Generator efficiency.
genhtval	Value of hot water from the generator cooling system.
gensize	The number of kilowatt-hours produced by the generator in 24 hours if operated at rated capacity.
hyalolim	The density of hyacinths remaining after harvest.
hyarea	The area of the hyacinth growth ponds.
hyenreq	The energy required to harvest hyacinths.
hyhrvfrq	The frequency at which hyacinths are harvested regardless of density.
hyhrvlim	The density at which hyacinths will be harvested, regardless of the day.
hylabreq	The amount of labor required to harvest hyacinths.

hysolids	The percent total solids of the hyacinth slurry fed to the CER.
hyvsofts	The pecent of the hyacinth total solids which are volatile solids.
intrstrt	The interest rate charged for capital costs of system construction.
laborcst	The cost of labor.
landcost	The cost of the land on which the system is constructed.
powercst	The cost of purchased electricity.
pwrvalue	The value of site-generated electricity.
rx1brc	FBR bacterial retention coefficient.
rx1hx	FBR heat exchanger efficiency.
rx1ktot	The number of compartments to simulate for the FBR.
rx1mint	The minimum temperature at which the FBR is to be operated.
rx1rval	The "R" value of the FBR insulation.
rx1vol	The liquid volume of the FBR.
rx2brc	CER bacterial retention coefficient.
rx2ktot	The number of compartments to simulate for the CER.
rx2maxvol	The maximum volume of the CER.
rx2sedvs	The VS concentration of the initial CER seed material.

rx2seed	The fraction of the CER volume left behind after emptying to act as a seed for the next digestion cycle.
rx3brc	CSTR bacterial retention coefficient.
rx3hx	CSTR heat exchanger efficiency.
rx3ktot	The number of compartments to simulate for the CSTR.
rx3mint	The minimum temperature at which the CSTR is to be operated.
rx3mxhrs	The number of hours per day the CSTR is mixed.
rx3rval	The "R" value of the CSTR insulation.
rx3vol	The liquid volume of the CSTR.
washteff	The efficiency of heat recovery by the generator cooling system.

APPENDIX C

PROGRAM FLOW CHART


Start
Input and Initialization
Event Handling
Flow Distribution
Reactor Simulation
FBR
CER
CSTR
Digestion
Algae Simulation
Hyacinth Simulation
Daily Summation
No
End
Yes
Economic Summary
Stop

BIOGRAPHICAL SKETCH

Michael V. Thomas was born on September 24, 1950, in South Haven, Michigan. He graduated from Niles senior High School in Niles, Michigan, in 1969. From 1969 to 1978 he served as an electronics technician in the U.S. Navy, earning an Associate of Arts degree from Florida Jr. College at Jacksonville while stationed at Jacksonville Naval Air Station. He entered the University of Florida in January 1979 and earned his B.S. in microbiology in 1981. In 1984 he was awarded a Master of Engineering degree in agricultural engineering from the University of Florida. He is married and has spent the last four years in pursuit of his doctoral degree in agricultural engineering.

I certify that I have read this study and that in my opinion it conforms to acceptable standards of scholarly presentation and is fully adequate, in scope and quality, as a dissertation for the degree of Doctor of Philosophy.

Roger A. Nordstedt

Roger A. Nordstedt, Chairman
Associate Professor of
Agricultural Engineering

I certify that I have read this study and that in my opinion it conforms to acceptable standards of scholarly presentation and is fully adequate, in scope and quality, as a dissertation for the degree of Doctor of Philosophy.

Arthur A. Teixeira

Arthur A. Teixeira, Cochairman
Associate Professor of
Agricultural Engineering

I certify that I have read this study and that in my opinion it conforms to acceptable standards of scholarly presentation and is fully adequate, in scope and quality, as a dissertation for the degree of Doctor of Philosophy.

Ben L. Koopman
Associate Professor of
Environmental Engineering
Sciences

I certify that I have read this study and that in my opinion it conforms to acceptable standards of scholarly presentation and is fully adequate, in scope and quality, as a dissertation for the degree of Doctor of Philosophy.

W S Otwell

Walter S. Otwell
Associate Professor of Food
Science and Human Nutrition

I certify that I have read this study and that in my opinion it conforms to acceptable standards of scholarly presentation and is fully adequate, in scope and quality, as a dissertation for the degree of Doctor of Philosophy.

J. Wayne Mishoe

Wayne Mishoe
Professor of Agricultural Engineering

This dissertation was submitted to the Graduate Faculty of the College of Engineering and to the Graduate School and was accepted as partial fulfillment of the requirements for the degree of Doctor of Philosophy.

December 1988

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